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Economic Analysis of Treatment Technologies to Achieve VOC (Volatile Organic Chemicals) Removal to Safe Levels

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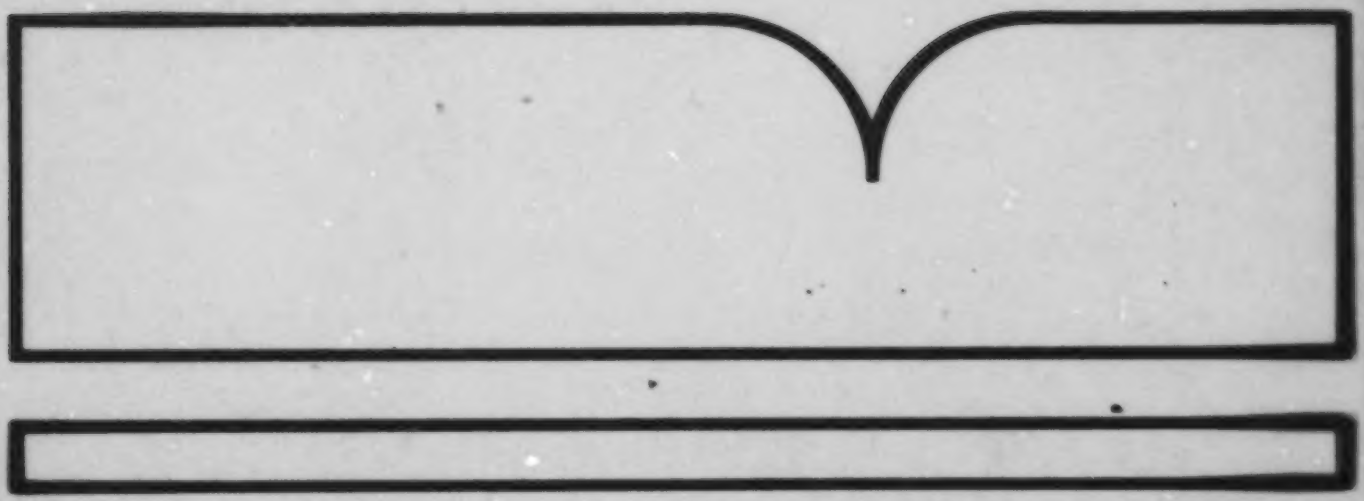

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16. ABSTRACT

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by

Robert M. Clark, Director
Drinking Water Research Division

Jeffrey Q. Adams, Environmental Engineer
Systems and Field Evaluation Branch
Drinking Water Research Division
Risk Reduction Engineering Laboratory
U.S. Environmental Protection Agency
Cincinnati, Ohio 45268

RISK REDUCTION ENGINEERING LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

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ECONOMIC ANALYSIS OF TREATMENT TECHNOLOGIES TO ACHIEVE VOC REMOVAL TO SAFE LEVELS

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INTRODUCTION

The Safe Drinking Water Act (SDWA) of 1974 was established as a preventive public health measure to assure the quality of drinking water for all U.S. consumers. It required USEPA to set regulations based on the 1962 U.S. Public Health Service Standards, to establish Maximum Contaminant Level Goals (MCLGs) and Maximum Contaminant Levels (MCLs) for contaminants with potential adverse health effects, and to protect groundwater by regulating underground waste injection and sole-source aquifers. The SDWA Amendment of 1986 represented another major a step in the regulation of drinking water quality in the U. S. The amendments establish an aggressive standard setting agenda and a short time frame in which the USEPA must revise existing standards and establish new ones.[1]

The SDWA amendments streamline USEPA's regulatory and enforcement procedures requiring extensive, substantive changes in the federal drinking water program. Nevertheless, although the pace of standard setting is greatly accelerated, the overall direction of national drinking water policy remains unchanged because many of the new standards represent regulations already under development by the USEPA.

The key elements of the 1986 amendments are:

- new standards for 83 contaminants by early June 1989 (as opposed to the current 25 standards);
- designation of best available technology (BAT) for each of the regulated contaminants;

- filtration of many surface water supplies;
- disinfection of all public water supplies;
- monitoring for contaminants that are not regulated;
- a ban on lead solders, flux, and pipe in public water systems; and a
- comprehensive groundwater protection program, particularly in areas around drinking water wells.

Under the new amendments, EPA will redesignate existing MCLs as primary regulations, and MCLGs will be set at the level "at which no known or anticipated adverse health effects of persons occur and which allows an adequate margin of safety." The MCLs will be set as close as "feasible" to the goal levels.

The new amendments also established the concept of BAT. The purposes of developing BAT are:

- to assist USEPA in setting drinking water standards, particularly by taking into consideration the capabilities of various treatment technologies for removing contaminants from water;
- to assist water utilities in selecting appropriate treatment methods to meet the regulations, considering the possibility of future changes in the drinking water standards;
- to identify treatment technologies that would be most appropriate for small systems; and
- to provide a basis for utilities to obtain a variance from meeting a given regulation.

In June of 1988 EPA set the first of eight standards for what are termed Volatile Organic Chemicals (VOC). Table 1 contains both the final MCLGs which are health based, and MCLs in micrograms per liter (ug/L) for these chemicals. Setting of these VOC limits has been called Phase I of the regulatory process.

The Drinking Water Research Division (DWRD) of EPA has a major role in determining BAT for a given contaminant or set of contaminants. The standard research protocol followed by DWRD is to evaluate unit processes at the bench level, test the process at the pilot scale and if its performance is promising, evaluate a prototype under field conditions.[2] At all stages, cost and performance are key factors in moving a technology to the next phase. Thus when a technology is tested at a field scale, cost and performance data are routinely collected.

In the VOC regulations both granular activated carbon (GAC) and packed tower aeration (PTA) have been designated as BAT. DWRD has performed a great deal of research both at the bench, pilot and field scale on the use of GAC and pilot and field scale research for packed tower aeration. Cost and performance data has been collected for both technologies. This paper will compare predictions of cost and performance of GAC and PTA using available models and field scale data. A sensitivity analysis of each technology with respect to design and operating variables and cost will be made. Liquid-phase GAC and PTA with off gas control will be compared. In the following sections each technology will be described, cost and performance will be discussed, and comparative evaluations will be made.

GRANULAR ACTIVATED CARBON

To date, the use of liquid-phase GAC for drinking water treatment in the U. S. has been limited to primarily taste and odor control applications.[2] However, since the widespread detection of synthetic organics in drinking water supplies, much research and many pilot-scale studies have been undertaken to evaluate the effectiveness of GAC for controlling organic compounds.

Based on past research and pilot-scale work, GAC represents one unit process with the ability to remove a broad spectrum of organic chemicals from water. Although GAC is considered to be the best available broad spectrum removal process, it exhibits a wide range of effectiveness in adsorbing organic compounds.

PROCESS DESCRIPTION

The application of granular activated carbon adsorption for removing organic compounds from drinking water supplies involves the following major process design considerations:

- Carbon usage rate - pounds of carbon per gallon of water treated
- Empty bed contact time
- Pretreatment
- Contactor configuration - downflow versus upflow, pressure versus gravity, single-stage versus multi-stage or parallel
- Method of GAC regneration - on-site versus off-site.

The carbon usage rate yields an indication of the rate at which carbon will be exhausted or replaced, and thus affects the cost of the treatment system. For a full-scale GAC installation the carbon usage rate often dictates the choice between providing an on-site regeneration system and replacing the spent carbon with virgin carbon. It also impacts any costs associated with carbon handling, such as storage, dewatering, attrition losses and transportation. The carbon usage rate for a given type of water and contaminant(s) can be estimated by different methods. These include:

- Isotherms and prediction models incorporating kinetics
- Dynamic mini-columns (DMC)
- Pilot-scale columns
- Full-scale columns.

The empty bed contact time (EBCT) provides an indication of the quantity of carbon which will be on-line at any one time, and thus impacts the capital cost for the system. The EBCT is an important design parameter and it has a significant impact on the carbon usage rate or bed life of the carbon in an adsorber for any particular compound or group of contaminants.

GAC systems may require some kind of pretreatment to prevent clogging of the carbon bed and to minimize the organic loading on the carbon. Clogging of the bed could be caused by suspended solids in the raw water or by precipitation of iron and manganese on the carbon. The former is typical of surface water systems. When the carbon bed life is long, clogging may also be caused by biological growths. Disinfection with chlorine prior to GAC adsorption should be avoided as chlorine is adsorbed by carbon, and, therefore, competes with the organics. A common solution to prevent clogging of the bed is to provide filtration ahead of the GAC system. Sometimes, GAC systems are added at the end of a conventional treatment process. In this mode, the GAC contactors are called post-filter adsorbers.

When the organic levels in the raw water are high, the carbon is exhausted at a fast rate, necessitating frequent replacement. This increases the operating cost of the system. Pretreatment can be provided to reduce the organic loading on the carbon and thereby increase the carbon bed life. The need for pretreatment should, however, be justified on the basis of costs. Examples of processes which may be used for pretreatment include conventional treatment, ozonation and packed column aeration.

Based on the estimates of carbon usage rate and contact time, a conceptual process design can be developed by evaluating various contactor configurations. The two basic modes of contactor operation are downflow and upflow. Upflow

beds typically have been applied to situations where very long contact times (greater than 120 minutes) are required and/or where the level of suspended solids is high. Downflow fixed bed contactors offer the simplest and most common contactor configuration for organic contaminant removal from drinking water. The contactors can be operated either under pressure or by gravity.

The choice of pressure or gravity is generally dependent upon the hydraulic constraints of a given system. Pressure contactors may be more applicable to ground water systems because pumping of the ground water is required. Gravity contactors are generally more suitable for surface water systems if sufficient head is available. Gravity contactors, when used, will typically be placed downstream of surface water filtration systems.

Another basic consideration in evaluating the design of a GAC system for contaminant removal is the method of carbon reactivation or replacement of spent carbon. The two basic approaches to regenerating the carbon are:

1. Off-site disposal or reactivation
2. On-site reactivation

Based on information from GAC manufacturers and conceptual design evaluations, on-site reactivation generally does not appear to be economical for systems where the carbon exhaustion rate is less than 1,500 pounds per day.

Adams et al. demonstrated that a reactivation facility having an operating capacity of 12,000 pounds of GAC per day could provide a cost-effective alternative to carbon replacement.[3] Moreover, utilization of the facility's excess capacity for a regional system showed that off-site reactivation would be more economical for the participating utilities than either carbon replacement or on-site reactivation.

Under the throwaway concept of off-site disposal, spent carbon is generally slurried by gravity to a draining bin where the free water is

removed and returned for treatment. The drained carbon is then manually drummed and shipped for landfill or incineration. Virgin GAC is replaced in the steel or concrete contactors. The advantages to this approach lie mainly in its technical simplicity; and, as such, it is a sound approach for applications requiring a relatively small carbon exhaustion rate (generally less than 500 pounds per day). The need to dispose of the spent carbon, however, is a definite drawback, especially since toxic or hazardous materials are adsorbed on the spent carbon. It may become necessary to consider incineration of the spent carbon to ensure proper ultimate disposal.

The off-site reactivation approach is somewhat similar from a carbon handling standpoint to the throwaway concept; however, it begins to assume some of the economies associated with carbon reuse. When compared to other alternatives, however, the number of handling steps and resulting carbon attrition and loss are a major disadvantage. The off-site reactivation approach has generally proven most cost effective in applications where the carbon exhaustion rate falls in the range of 500 to 2,000 pound per day.[4]

TREATABILITY STUDIES

Studies on the treatability of the VOCs can be classified into four types of evaluations: isotherms, bench-scale dynamic minicolumns (DMC), pilot scale columns and full scale columns. Isotherm evaluations are batch tests which yield the equilibrium or maximum VOC loading on a particular carbon at a given VOC concentration. Bench scale tests involve the use of dynamic mini-columns to estimate carbon usage rates under flow-through conditions. Pilot tests are conducted using larger columns than are used in mini-column testing and thus require significantly greater quantities of water and time. Full scale test evaluates the performance of GAC in actual field installations.

Adsorption isotherms are useful screening tools for determining preliminary carbon usage rates, and evaluating the relative adsorbability of a particular compound in comparison to other compounds. The analytical procedure that is generally followed in conducting isotherm tests is outlined by Randtke and Snoeyink.[5] The procedure involves placing a measured weight of pulverized carbon in water/VOC solution of known VOC concentraton and agitating over a sufficient time to reach equilibrium. The resultant effluent VOC concentration is then measured and equilibrium capacity (or loading) calculated from the amount of VOC adsorbed and the known weight of carbon in solution. Equilibrium capacities are determined at several effluent concentrations for each VOC. The relationship between equilibrium capacity and effluent concentration has been found to generally follow the Freundlich isotherm relationship:

$$X/M = K\ C^{1/n} \qquad (1)$$

where:

X/M = equilibrium capacity (mg VOC/g carbon)
K = capacity at 1 mg/L VOC concentration
C = VOC equilibrium concentration (mg/L)
1/n = exponent

K and 1/n are typically referred to as Freundlich constants.

AERATION

Air stripping has been used effectively in water treatment to reduce the concentration of taste and odor producing compounds and certain organic compounds.[2] Aeration, or air stripping, may be described as the transfer of a substance from liquid phase to gas phase. The driving force for mass transfer is the difference between actual conditions in the air stripping unit and conditions associated with equilibrium between the gas and liquid

phases. The equilibrium concentration of a solute in air is directly proportional to the concentration of the solute in water at a given temperature, according to Henry's Law which states that the amount of gas that dissolves in a given quantity of liquid, at constant temperature and total pressure, is directly proportional to the partial pressure of the gas above the solution. Thus, the Henry's coefficients describe the relative tendency for a compound to separate between gas and liquid. Henry's coefficients can be used to give a preliminary indication of how well a VOC can be removed from water via air stripping.

Henry's coefficients for selected VOCs are presented in Table 1. The magnitudes of the constants for the various compounds are functions of their solubility in the liquid phase and their volatility. A high Henry's coefficient indicates equilibrium favoring the gaseous phase; i.e., the compound generally is more easily stripped from water than one with a lower Henry's coefficient. Theoretical Henry's coefficients can be estimated from vapor pressure, solubility and molecular weight as follows:

$$H = \frac{V_p \times MV}{sol} \times 73.1 \qquad (2)$$

where: H = Henry's coefficient (atm)

V_p = vapor pressure (mm Hg)

MW = molecular weight (g m^{-1})

sol = solubility (mg L^{-1})

The mass transfer coefficient relates the driving force (concentration gradient) to the actual quantity of material transferred from liquid to air. The mass transfer coefficient is a function of the physical/chemical properties of an individual VOC, the type of packing material used, and the

gas and liquid loading rates. In packed columns, packing materials are used which provide high void volumes and high surface area. The water flows downward by gravity and air is forced upward. The untreated water is usually distributed on the top of the packing with distribution trays and the air is moved through the tower by forced or induced draft. This design results in continuous and thorough contact of the liquid with the gas and minimizes the thickness of the water layer on the packing, thus promoting efficient mass transfer.

The design of air stripping equipment has been extensively developed in the chemical engineering industry for handling concentrated organic solutions. The procedures used in the chemical engineering literature can also be applied to water treatment applications for trace organics removal. The rate at which a volatile compound is removed from water by aeration depends on several factors:

- Air:water ratio
- Contact time
- Available area for mass transfer
- Temperature of the water and the air
- Physical and chemical properties of the contaminant.

The first three factors may be controlled in the design of an air stripping unit, while the other two factors are set for a specific water supply.

The air flow requirements for a packed column depend on the Henry's coefficient for the particular compound(s) to be removed from the water. In an ideal aeration system, the minimum air:water ratio which will achieve complete removal of a contaminant is proportional to the reciprocal of the Henry's coefficient. The greater the Henry's coefficient, the less air that is required to remove the compound from water. Because aeration systems are not ideal, actual air:water ratios to achieve a given removal efficiency are greater than the theoretical minimum air:water ratios. The contact

time is a function of the depth of the packing material. An increase in the depth of packing material results in a greater contact time between the air and the water, and consequently, will result in higher VOC removals.

The available area for mass transfer is a function of the packing material. Various sizes and types of packing material are available including 1/4 inch to 3-inch sizes in metal, ceramic and plastic materials. In general, the smaller packing material provide a greater available area for mass transfer per volume of material, thus increasing the mass of contaminant removed. However, smaller packing sizes increase the pressure drop through the air stripping tower.

COST MODEL DEVELOPMENT

EPA's DWRD has conducted a number of research efforts focused on the economic aspects of various treatment technologies. These data have been updated and refined to provide a preliminary assessment of the cost impact of the regulations to drinking water utilities.

Studies by Gumerman and Culp, et. al. included cost estimates for various treatment scenarios.[6,7] As-built designs and standard cost reference documents were used to determine the amount of excavation, framework, and materials such as concrete and steel required. Information gathered from existing plants and manufacturers were used to determine the cost of equipment, supplies, and operating requirements. The construction and operating costs for each unit process were presented as a function of the process design or operating parameters that were determined to be the most useful and flexible under varying conditions. For example, GAC post-filter adsorber construction costs were presented as a function of the effective GAC volume (cu. ft.) in the

contactors, an approach that allows various EBCTs to be examined. This approach provided more opportunities for analysis of cost sensitivities than if the cost data were related only to water flowrate in the treatment plant.

Construction cost estimates were determined by aggregating the costs of several components: excavation and sitework, manufactured equipment, concrete, steel, pipes and valves, electrical and instrumentation, labor, and building. The construction cost estimates excluded engineering and contractor fees. These fees and other special contingencies were required to be added seperately in order to provide a total capital cost estimate. Operation and maintenance (O&M) requirements were developed for unit process energy and fuel, building-related energy, maintenance and operating materials, and labor. Provisions for updating cost values to account for inflation over time were made using several component indicies or general indicies such as the Engineering News Record (ENR) Construction Cost Index and Producers Price Index.

Parallel to the development of the conceptual design cost data, DWRD was conducting studies designed to collect specific cost and performance data on full-scale GAC and aeration treatment at several field sites. GAC steel pressure contactors used as post filter adsorbers were studied along with GAC adsorbers converted from conventional sand filters at Cincinnati, Ohio.[8] On-site carbon reactivation using a fluid-bed furnace was also evaluated. Onsite GAC reactivation and regional reactivation were studied at Manchester, New Hampshire.[9] Studies at Jefferson Parish, Louisiana included the examination of GAC pressure contactors and on-site reactivation with an infrared furnace.[10] Pilot and field scale aeration studies have been conducted at Wausau, Wisconsin; Baldwin Park, California; and Brewster, New York.[2]

Data from conceptual design studies and field-scale research efforts have been used to develop preliminary cost estimating functions, as described by Clark and Adams.[11,12] The cost functions can be used to examine the sensitivity of treatment costs to variations in system design and operation, and economic parameters. The cost analysis presented herein builds upon the cost estimating approach developed by Clark and Adams.

COST AND PERFORMANCE OF GAC TREATMENT SYSTEMS

For purposes of the GAC cost analysis presented herein, the constant-pattern-homogeneous-surface-diffusion model (CPHSDM) was used to predict liquid-phase GAC use rates for selected single-solute VOCs. A discussion of the assumptions and application of this model is given by Hand et al.[13] Input parameters required by the CPHSDM are: the Freundlich K and 1/n values as determined from an isotherm, the surface diffusion and film transfer coefficients, mean carbon particle radius, the bulk and apparent particle densities, superficial fluid velocity, empty bed contact time (EBCT), and the initial concentration of the VOC. Where surface diffusion and film transfer coefficients were not available, literature correlations were employed.

Table 2 presents for each VOC examined herein, the CPHSDM-predicted bed lives associated with EBCTs of 10 and 15 minutes, given the specified influent and effluent concentration criteria, single-solute isotherm data, and type of water in which studies were conducted. Predictions were based upon the assumptions of a single full-scale contactor, employing Filtrasorb 400 GAC, and a hydraulic loading of 4 gpm/ft^2 (9.6 m/hr).

Only one of the compounds, p-dichlorobenzene, exhibited a bed life greater than 2 years, typical of taste-and-odor control applications using GAC. Most of the VOCs exhibited bed lives in the range of 6 to 24 months

with EBCTs of 10 and 15 minutes. Three VOCs (1,2-dichloroethane; 1,1,1-trichloroethylene; and vinyl chloride) had predicted bed lives of less than 3 months. For these relatively moderate and poorly adsorbing compounds, the limitations of the CPHSDM predicted use rates are of concern. The data presented in Table 2 involve single-solute VOCs. In natural waters, multi-solute VOCs would be expected. A poorly-adsorbed VOC in the presence of more strongly adsorbed VOCs would likely exhibit higher GAC use rates (shorter bed lives) resulting from competitive adsorption and displacement. Even strongly-adsorbed VOCs in the presence of other strongly-adsorbed VOCs may exhibit higher use rates resulting from competitive adsorption. Further, the modeling used in this analysis did not account for possible reduction of the GAC's capacity as a result of background organic matter being preadsorbed on carbon. The affects of backwashing the carbon were also not considered.

In spite of the limitations of the CPHSDM approach, the predicted use rates for the selected VOCs give a general picture of adsorbability, useful for preliminary cost analysis. The development and application of cost data is presented below, and is integrated with the predicted use rate (bed-life) data to provide cost estimates for controlling the selected VOCs by GAC treatment.

Preliminary total cost estimates for liquid-phase GAC treatment systems have been developed for a range of plant sizes, EBCTs and carbon bed lives. Table 3 presents economic parameters used in this analysis. The GAC systems analyzed were assumed to treat an average flow of water equal to 70% of the system's design capacity. GAC contactors were assumed to be in a parallel configuration and post-filter mode. Individual contactors were assumed to operate at design hydraulic loading whenever possible. For systems that

were less than 10 MGD in size, steel pressure GAC adsorbers were employed, otherwise, concrete gravity contactors were assumed. When spent carbon requirements were less than 1500 pounds/day (680 kg/day) then replacement of spent carbon with virgin GAC was assumed. When carbon requirements were greater than 1500 pounds/day then multihearth on-site reactivation was employed. GAC reactivators were designed to have excess capacity to allow for 25% operational downtime. Individual reactivators were limited to a design capacity of 50,000 pounds/day (22,680 kg/day).

Tables 4 and 5 present total cost estimates (¢/1000 gal) for GAC treatment systems with EBCTs of 10 and 15 minutes, respectively. The contrast between the high unit costs associated with very small systems having short carbon bed lives relative to the low unit costs associated with large systems having long GAC bed lives is clearly seen from these tables. Figure 1 shows GAC treatment cost estimating curves as a function of system design capacity associated with GAC bed lives of 3, 6, and 12 month periods. An EBCT of 10 minutes is assumed. Economies of scale are such that a steep drop in unit cost is exhibited with increasing system size up to about 30 MGD, and the cost decreases gradually, thereafter.

Figure 2 presents preliminary GAC treatment cost estimating curves associated with system sizes of 1, 10 and 100 MGD as a function of carbon bed life. An EBCT of 10 minutes is assumed. The cost curves are relatively steep with respect to GAC bed life in the range up to about 5 months, and they level out as bed life increases. For those VOCs with influent and effluent criteria resulting in carbon bed lives shorter than 5 months, accuracy in determining GAC use rates becomes a more critical issue. The shorter the carbon bed life, the more sensitive the cost becomes to variations in the GAC use rate.

Figure 2 shows where various VOCs with their specified criteria fall on the cost curves, based on the results of the adsorption modeling presented earlier. Most of the VOCs presented in Table 2 exhibited bed lives of 6 months or longer, falling on the fairly-flat portion of the cost curves. Cost estimates for these scenarios range from about 34-45 ¢/1000 gal for a 1 MGD system, to about 20-30 ¢/1000 gal for a 10 MGD system, to about 11-15 ¢/1000 gal for a 100 MGD system. Inaccuracy in determining GAC use rates may not be a significant cost issue for the moderate and strongly adsorbing VOCs. Because the GAC use rates used in this analysis were based on single-solute VOCs, a reduction in bed life is expected in actual performance as a result of competing VOCs in the water and preadsorption of background organic matter on the carbon. However, even if bed lives are reduced by 50%, the relative increase in unit costs for the more strongly adsorbed VOCs do not appear to be very significant, especially for larger systems.

The GAC cost estimates for controlling the relatively poorer adsorbing VOCs like vinyl chloride, 1,2-dichloroethane, and 1,1,1-trichloroethane at high influent concentrations are subject to greater use rate and cost variability, especially for small sized systems. For these more difficult to adsorb compounds, Henry's coefficients are high enough that they may be more economically controlled using packed tower aeration or a combination of aeration and GAC.

Even though the predicted use rates for VOCs in this analysis do not take into account some of the factors that may reduce the bed life of carbon in actual performance, they also do not account for some factors that may extend the life of the carbon. In full-scale operation, biological activity may increase the GAC bed life by degrading the VOCs of interest. In this analysis, the use rate predictions were based on individual contactors operating in a parallel configuration. In many cases, contactors designed

for series operations may extend the bed life of the carbon. Dyksen (1987)[15] and Zimmer (1987)[16] cite examples comparing systems with two adsorbers in series versus single adsorber operations having the same total EBCT. The in-series systems were predicted to treat about 40% more water.

In addition to these factors, the simplified use rate analysis in this paper did not consider reactivation/replacement schedules based on monitoring the blended effluent from multiple adsorbers instead of the effluent from each individual contactor. By blending the effluent of two or more contactors operating on a staggered breakthrough schedule, the bed lives of individual contactors can be extended. An example of this approach is discussed by McGuire, et al. (1988) concerning GAC treatment for the Metropolitan Water District of Southern California.[17]

The potential increase in carbon bed life from the utilization of a series adsorber scheme and/or blending the effluent from two or more contactors will partially offset the reduction in bed life resulting from competition among multi-solute compounds in natural waters and preadsorption of other background organic matter on the carbon. Another operational strategy that may be considered to reduce treatment costs for small systems is the utilization of regional or commerical off-site carbon reactivation instead of on-site facilities or complete virgin GAC replacement. Regional reactivation was successfully employed at Manchester, New Hampshire.[9]

COST AND PERFORMANCE OF PACKED TOWER AERATION

For purposes of the cost analysis presented herein, a performance design and analysis model for packed tower aeration (PTA), based on work by Hand et al., was utilized.[14] A computer program has been developed by the DWRD to integrate cost and performance models, to provide preliminary estimates for controlling VOCs by PTA. The approach requires one to assume values for two design variables in each scenario, and analyze a number of cases to find cost effective PTA designs. For example, given system design and operating flows, treating specified VOCs with influent and effluent concentrations, an air: water ratio or stripping factor [air:water ratio * Henry's coefficient] is assumed, along with a value for pressure drop gradient thru the packing media. Model outputs include: tower packing media depth required for the specified level of treatment, tower diameter, loading rate, power required for blowers and pumps, and cost estimates for capital and operation and maintenance.

Figure 3 shows the design tradeoffs between tower packing media volume and tower electric power usage as a function of air:water ratio and pressure drop gradient thru the packing media. This case assumes a 1 mgd system treating 95% removal of 1,2-dichloroethane. For a given packing media height, increasing the air:water ratio increases pressure drop thru the tower. From a design standpoint, given a fixed air:water ratio, increasing the assumed pressure drop increases the required packing depth but decreases the tower diameter enough to result in a lower overall packing volume. For example, given an air:water ratio of 50, the packing depth increases from 68.9 to 73.3 feet as pressure drop increases from 50 to 100 $N/m^2/m$, but effective diameter decreases from 6.6 to 5.4 feet, resulting in a decrease in volume from 2360 ft^3 to 1690 ft^3. Smaller packing volumes result in lower capital costs.

However, because packing height increases with increasing pressure drop, electrical power usage (pumps and blowers) increases as shown in Figure 3. From another design standpoint, assuming a fixed pressure drop thru the media, increasing the air:water ratio decreases the required packing height. This results in a lower packing volume and lower electrical power requirements for pumping. However, an increase in air:water ratio increases the electrical power requirements for blowers. This is especially significant when the air:water ratio is beyond the optimal design. Therefore, variations in air:water ratio and pressure drop gradient thru the media result in cost tradeoffs between capital and electrical power requirements.

Economic and system parameters assumed in the PTA analysis are given in Table 6. Figure 4 shows the sensitivity of cost to variations in two key design variables, air:water ratio and pressure drop thru the packing media. In this example, a PTA system with a design capacity of 1 mgd (0.7 mgd operating flow) is assumed for controlling 1,2-dichloroethane (1,2-DCA) at 95% removal. The total unit cost (¢/1000 gallons) of the system decreases from 29 to 21 ¢/1000 gal as air:water ratio increases from 50 to 90, over a pressure drop range of 50 to 70 $N/m^2/m$. At air:water ratios greater than 90, the unit cost begins to increase. At the optimal air:water ratio of 90, the cost varies by as much as 5 ¢/1000 gal over a pressure drop range of 50 to 190 $N/m^2/m$. The tower packing depth required decreases from 70 to 29 feet as air:water increases from 50 to 90. Both estimated system cost and tower height are important factors in selecting practical PTA designs under site specific conditions.

The Henry's coefficient of a compound has a significant impact on the cost effective design of a PTA system. Figure 4 presented an example for

controlling 1,2-DCA with a Henry's coefficient of 0.023 m^3 water/m^3 air. Figure 5 shows a similar analysis of cost variation as a function of air: water ratio and pressure gradient for vinyl chloride at 98% removal with a Henry's coefficient of 2.4. The higher the Henry's coefficient, the more easily the VOC can be stripped, requiring lower optimal air:water ratio and packing depth. For vinyl chloride, the optimal air:water ratio appears to be about 10 with a packing depth of 18 feet, resulting in a unit cost of about 11 ¢/1000 gal. However, for this VOC having a high Henry's coefficient, cost is very insensitive to changes in both air:water ratio and pressure drop over the ranges shown. The cost varies by only 2 ¢/1000 gal (from 13 to 11 ¢/1000 gal) over the range of air:water ratio from 1 to 50, and very little cost variation with respect to changes in pressure gradient.

The required percent removal of a compound has an important impact on the cost effective design of a PTA system. Figure 6 presents cost estimating curves as a function of percent removal for three VOCs; 1,2-DCA, trichloroethylene (TCE), and vinyl chloride, which span the range of Henry's coefficients and stripability examined herein. The optimal air:water ratio, packing depth, and resulting unit cost are more sensitive to increases in percent removal for those VOCs with relatively lower Henry's coefficients, especially in the range of 90 to 99 percent removal. For the most easily stripped VOC, vinyl chloride, increases in percent removal have little impact on the required air:water ratio, packing depth, and resulting cost, except at removal levels nearing 100%.

The Henry's coefficient of a compound has the greatest impact on the cost effective design of a PTA system. The optimal selection of air:water ratio, pressure drop thru the media, packing depth, and resulting cost are most dependent on the Henry's coefficient. Figure 7 presents cost estimates

for the eight VOCs examined herein for 95% removal at three system sizes. The system unit cost is presented as a function of Henry's coefficient, and the cost minimized air:water ratio, stripping factor, and packing depth are given. As Henry's coefficient increases, system cost decreases, optimal air:water ratio and packing depth decrease, and optimal stripping factor appears to increase. For this example at 95% removal, the unit cost approximately doubles from a Henry's coefficient of 2 to 0.02 m^3 water/m^3 air; optimal air:water ratio increases from 10 to 90 and required packing depth increases from about 14 to 29 feet over this range. The cost curves flatten out with increasing Henry's coefficient at about 0.5, resulting in only a small decrease in unit cost with higher Henry's coefficient values.

As mentioned earlier, the type and size of packing material can affect the cost and performance of a PTA system. Smaller packing material provides greater area for mass transfer per volume of material, and may allow for a lower air:water ratio and shorter tower. For example, the previous PTA analysis assumes the use of 2-inch plastic saddles for packing material. The minimized cost associated with 95% removal of 1,2-DCA for a 1 mgd system is 21.2 ¢/1000 gal with an air:water ratio of 90 and packing depth of 29 feet. If 3.5-inch plastic tripacks are used for packing material, then the estimated cost is 25.6 ¢/1000 gal with an air:water ratio of 98 and packing depth of 42 feet.

Water temperature can also affect the performance and cost of PTA systems. For example, the prior analysis assumes a water temperature of 12°C. the cost shown in the previous example was 21.2 ¢/1000 gal, with packing depth of 29 feet. If the temperature is 20°C then the estimated cost decreases to 19 ¢/1000 gal, with packing depth of 23 feet.

PTA WITH OFF-GAS CONTROL

The previous analysis assumes no treatment of the air stripper off-gas. Recently, there has been a great deal of concern expressed regarding the problems of "cleaning up" air emissions from PTA systems. In cases where air-phase GAC for tower off-gas control is required, the cost of air stripping could more than double, making liquid-phase GAC a competitive control technology.

Cost estimates for PTA systems with gas-phase GAC treatment have been developed using a design and analysis approach based on work by Crittenden, et al.[18] Gas-phase GAC usage rates were determined from the Dubinin-Radushkevich (D-R) equation. Off-gas control costs include capital and operation and maintenance requirements for blowers, heaters, gas-phase GAC adsorbers, and off-site carbon regeneration. Costs for heating the tower off-gas from 12 to 24°C were included to reduce the relative humidity to under 40%, thus minimizing the affects of humidity on the carbon's adsorption capacity.

Figure 7 shows the cost minimized air:water ratio and packing depth associated with controlling each VOC examined, assuming no tower off-gas control. When gas-phase GAC treatment is included, a lower air:water ratio and taller tower may be more economical. Increased air flow rates require gas-phase GAC adsorbers with greater cross-sectional area and increased carbon quantities, thus increasing capital costs. However, a lower air:water ratio requires a greater stripping tower packing depth which offsets some of the capital cost savings from the gas-phase GAC adsorbers. For example, a 1-mgd system controlling 1,2-DCA at 95% removal (Co=100, Ce=5 ug/L) assuming no off-gas control, exhibited a minimized cost of 21.2 ¢/1000 gal with an air:water ratio of 90 and packing depth of 29 feet. If the same system includes gas-phase GAC treatment, the initial cost effective air:water ratio is reduced

from 90 to 60. A total cost of 55.5 ¢/1000 gal is associated with an air:water ratio of 60 versus 65.5 ¢/1000 gal with an air:water ratio of 90. However, with air:water ratio of 60, the packing depth required is 45.4 feet compared to 29 feet with an air:water ratio of 90. In this case, a packing depth of 45 feet may be an impractical design for a single tower. Using two towers in-series way be a solution, but this will increase costs significantly. Systems must be designed and analyzed on a case by case basis, taking into account site specific conditions.

Figure 6 showed PTA unit cost as a function of percent removal, assuming no tower off-gas control. The cost of treatment is the same for a given removal percentage regardless of the influent concentration. However, when gas-phase GAC adsorption is included in the tower design then influent concentration of the VOC is important. For example, the 1 mgd PTA system with off-gas control treating 1,2-DCA at Co=100 ug/L, Ce=5 ug/L (95% removal) exhibits a cost of 65.5 ¢/1000 gal. The same system treating Co=500 ug/L, Ce=25 ug/L (95% removal) has a cost of 73 ¢/1000 gal. The higher influent concentration increases the gas-phase GAC exhaustion rate, requiring more frequent carbon regeneration, resulting in higher costs.

Figure 8 presents PTA unit cost estimating curves for 95 and 99% removal (Co=100 ug/L) with and without off-gas control, associated with a 1 mgd system, as a function of Henry's coefficient. System cost approximately doubles or triples when gas-phase GAC treatment is included in the PTA system. System cost increased by about 10 to 30% when VOC removal is increased from 95 to 99%. Unit costs for PTA systems with off-gas control show a steep drop over the Henry's coefficient range of 0.02 to 0.2 m^3 water/m^3 air). For VOCs with a Henry's coefficient in this range, the gas-phase GAC usage rate decreases resulting in less frequent carbon regeneration and lower costs. Also, as

Henry's coefficient decreases, the air:water ratio required decreases. A lower air flow rate allows for a smaller cross-sectional area needed for the GAC adsorbers and smaller quantities of GAC, resulting in lower capital costs. A steep increase in unit cost is exhibited for systems with off-gas control treating VOCs with a Henry's coefficient higher than about 0.2, because of a significant increase in the gas-phase carbon exhaustion rate. Tables 7 and 8 present cost estimates associated with four PTA system sizes, with and without off-gas control, for the eight VOCs examined, at 95 and 99% removal, respectively. Also, given are the cost minimized air:water ratio, packing depth, and stripping factor in each case.

COMPARISON OF LIQUID-PHASE GAC AND PTA TREATMENT

Having discussed liquid-phase GAC and PTA treatment separately, a comparison of the two technologies is presented for the eight VOCs examined. Figures 9 thru 16 show preliminary cost tradeoffs between liquid-phase GAC and PTA treatment as a function of system size from 1 mgd to 100 mgd. Figure 9 concerns the treatment of 1,2-dichloroethane from an influent concentration (Co) of 100 ug/L to an effluent (Ce) of 5 ug/L. Liquid-phase GAC treatment assumes an EBCT of 10 minutes. PTA treatment, in the case of excluding tower off-gas control, has significantly lower cost estimates than liquid-phase GAC systems. This remains true for all eight VOCs examined in this analysis. In this case, liquid-phase GAC treatment is more than twice as expensive as PTA without off-gas control. However, if tower off-gas control is required, then the PTA cost more than triples, and for systems larger than about 10 mgd, liquid-phase GAC appears to be more cost effective.

Figure 10 presents PTA and GAC (EBCT = 10 minutes) cost estimates for controlling p-dichlorobenzene from Co = 500 ug/L to Ce = 75 ug/L. If tower off-gas control is required for PTA treatment, then liquid-phase GAC appears to be significantly more cost effective, especially for larger sized systems.

PTA and GAC cost estimates for treating benzene from Co = 100 ug/L to Ce = 5 ug/L are presented in Figure 11. If tower off-gas control is included then PTA and GAC exhibit about the same cost for very small systems. As system size increases, GAC treatment appears to be more economical. Figure 12 shows cost estimates associated with reducing trichloroethylene from 100 to 5 ug/L. If off-gas control is included with the stripping tower then PTA exhibits a lower cost than liquid-phase GAC for systems under 30 mgd in size. For larger systems, GAC treatment appears to be cost effective.

Figures 13 and 14 show that for controlling 1,1,1-trichloroethane and carbon tetrachloride, PTA including off-gas control is more economical than liquid-phase GAC treatment, especially with smaller sized systems. Figure 15 presents cost estimates for treating 1,1-dichloroethylene from Co = 100 ug/L to Ce = 7 ug/L. Cost tradeoffs similar to the cases with controlling 1,2-dichloroethane and trichloroethylene are exhibited here. PTA with off-gas control appears to be cost effective for systems under 10 mgd in size, and GAC appears to be more economical for larger systems.

Cost tradeoffs between PTA without off-gas control and liquid-phase GAC, for the treatment of vinyl chloride from Co = 100 ug/L to Ce = 2 ug/L, are presented in Figure 16. The cost of PTA including off-gas control is not shown because it exceeds the range of cost values shown. If tower off-gas control is not required then PTA treatment is relatively inexpensive. However, if air emissions are a concern, then liquid-phase GAC treatment, though

relatively expensive, is probably the cheaper alternative because of the high costs of PTA off-gas treatment.

SUMMARY AND CONCLUSIONS

Both granular activated carbon (GAC) and packed tower aeration (PTA) have been designated as best available treatment technology in the VOC regulations. Cost and performance models have been developed to examine various treatment scenarios for controlling VOCs.

The constant-pattern-homogeneous-surface-diffusion model (CPHSDM) was utilized to predict liquid-phase GAC use rates for selected single-solute VOCs. In this analysis, eight currently regulated VOCs were examined. Only one of the compounds, p-dichlorobenzene, exhibited a bed life greater than 2 years, typical of taste-and-odor control applications using GAC. Most of the VOCs exhibited bed lives in the the range of 6 to 24 months with EBCTs of 10 and 15 minutes. Three VOCs (1,2-dichloroethane; 1,1,1-trichloroethylene; and vinyl chloride) had predicted bed lives of less than 3 months. For these relatively moderate and poorly adsorbing compounds, the limitations of the CPHSDM predicted use rates are of concern. In spite of the limitations of the CPHSDM approach, the predicted GAC bed-lives give a general picture of adsorbability, useful for preliminary cost analysis.

Preliminary cost estimates for liquid-phase GAC treatment systems have been developed for a range of plant sizes, EBCTs and carbon bed lives. Most of VOCs examined exhibited bed lives of 6 months or longer. Cost estimates for these scenarios range from about 34-45 ¢/1000 gal for a 1 mgd system, to about 20-30 ¢/1000 gal for a 10 mgd system, to about 11-15 ¢/1000 gal for a 100 mgd system.

A cost and performance model has been developed to examine various scenarios for controlling VOCs by packed tower aeration including off-gas treatment. Cost minimized designs were determined by varying parameters such as air:water ratio and pressure drop thru the packing media. The affects of influent and effluent concentrations, design and operating flows, Henry's coefficient, water and air temperatures, and type and size of packing material were examined. Preliminary cost estimates for PTA excluding off-gas control varied from about 49 ¢/1000 gal (treating 1,2-dichloroethane) to 35 ¢/1000 gal (treating vinyl chloride) for a 0.1 mgd system, 21 to 10 ¢/1000 gal for a 1 mgd system, and 14 to 6 ¢/1000 gal for a 10 mgd system. System cost approximately doubles or triples when gas-phase GAC treatment is included in the PTA system for off-gas control.

A comparison was made between liquid-phase GAC and PTA treatment alternatives. For all VOCs in this analysis, PTA treatment was more cost effective than liquid-phase GAC at all system sizes when assuming no PTA off-gas control. When PTA off-gas control was included, cost tradeoffs between PTA and GAC were seen at various systems sizes. Liquid-phase GAC appeared to be more cost effective than PTA with off-gas control at system sizes greater than 10 to 30 mgd, when treating the following VOCs: 1,2-dichloroethane; para-dichlorobenzene; benzene; trichloroethylene; 1,1-dichloroethene; and vinyl chloride. PTA treatment with off-gas control appeared to be more economical than liquid-phase GAC for treating 1,1,1-trichloroethane and carbon tetrachloride, especially with small systems.

This analysis provides insights into the sensitivity of GAC and PTA costs to variations in design and operating parameters. However, the evaluation of alternative treatment technologies must be made on a case by case basis, taking into account site specific conditions.

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Table 1. VOCs: MCLGs, MCLs and Henry's Coefficients

	(ug/L)[1]		Henry's Coefficient (atm m^3 m^{-3})	
	MCLG	MCL	[2]	[3]
1,2-Dichloroethane	0	5	0.027	0.023
para-Dichlorobenzene	75	75	0.042	-
Benzene	0	5	0.06	0.106
Trichloroethylene	0	5	0.125	0.116
1,1,1-Trichloroethane	200	200	0.23	0.172
Carbon Tetrachloride	0	5	0.21	0.556
1,1-Dichloroethylene	7	7	0.4	0.935
Vinyl Chloride	0	2	2.4	265

1. Final MCLGs were published Nov. 13, 1985. The MCLG and MCL for p-dichlorobenzene were reproposed at zero and 5 ug/L on April 17, 1987; comment was requested on levels of 75 ug/L and 75 ug/L, respectively.

2. Source: Cummins, M., Westrick, J. (1987) "Feasibility of Air Stripping for Controlling Moderately Volatile Synthetic Organic Chemicals", AWWA Annual Conference Proceedings, Denver, CO.

3. Source: Crittenden, J. et al. (1988) "Using GAC to Remove VOCs from Air Stripper Off-Gas", Journal AWWA, Vol 80, No. 5, May 1988.

Table 2. Full-Scale CPHSDM Predicted GAC Performance for Single Solute VOCs

	Water Type	Freundlich[d] Isotherm K[b]	1/n	Selected Criteria (ug/l) Co	Ce	CPHSDM[a] Predicted Bed Lives (days) EBCT = 10 min	EBCT = 15 min
1,2-Dichloroethane	DD	15.0	0.533	100	5	45	91
para-Dichlorobenzene	DD	1063.0	0.691	500	75	2123	3375
Benzene	DD	164.6	0.533	100	5	441	693
Trichloroethylene	OR	95.3	0.484	100	5	340	529
1,1,1-Trichloroethane	OR	29.7	0.495	500	200	50	77
Carbon Tetrachloride	DD	50.1	0.594	100	5	178	280
1,1-Dichloroethylene	DD	51.1	0.515	100	7	158	245
Vinyl Chloride	DD	3.6[c]	0.83	100	2	7	12

Note: [a]Single contactor, 4 gpm/ft^2, Filtrasorb 400, PSDFR = 3.0, 24°C
[b]K in (umol/g) (L/umol)** 1/n
[c]Estimated using Polanyi adsorption potential theory
[d]Source: Speth, T., Miltner, R., Crittenden, J., and Hand, D., "An Evaluation of GAC Adsorption Capacities for SOCs", USEPA, RREL, DWRD, 1988.

Water type: DD = distilled/deionized
OR = filtered Ohio River

Table 3. Parameters used in Liquid-Phase GAC Analysis

GAC Contactor System Operated at 70% of Capacity
Reactivators Designed for 75% Up-time
GAC Reactivation Losses at 12%
Capital Amortized at 10% Over 20 Years
GAC Price at 0.90 $/Pound at 100,000 Pounds
Labor and Fringe Rate at 15 $/hour
Electric Rate at 0.08 $/kwh
Fuel Oil at 0.95 $/Gallon
Process Water at 0.35 $/1000 Gallons
Contractor Profit Factor 5%
Special Sitework Factor 5%
Engineering Fees Factor 10%
Construction Contingencies Factor 10%
ENR Construction Cost Index 421 (June 1988)
Producers Price Index 299 (May 1988)

Table 4. GAC Treatment Cost Estimates by System Size and Bed Life; EBCT = 10 Minutes

GAC Bed Life (days)	System Cost(cents/1000 gallons)[a] Design Size (MGD)													
	0.1	0.25	0.5	1	2.5	5	10	15	25	50	75	100	150	200
30	203.0	148.7	127.4	111.6	96.8	71.2	56.8	48.3	41.8	35.3	35.0	33.6	32.9	32.3
45	174.2	121.0	100.4	85.3	76.7	60.5	47.8	40.0	34.2	28.5	27.0	27.0	25.0	24.6
60	159.6	106.9	86.7	72.0	63.8	56.0	42.8	35.5	30.2	25.0	22.6	21.3	21.3	20.0
75	150.8	98.4	78.4	64.0	56.1	51.1	39.7	32.6	27.6	22.7	20.5	19.3	19.3	18.1
90	144.8	92.7	72.8	58.5	50.8	46.0	37.5	30.6	25.9	21.2	19.1	17.8	16.7	16.7
105	140.6	88.6	68.8	54.6	47.1	42.3	35.9	29.2	24.5	20.1	18.0	16.8	15.4	14.5
122	137.0	85.1	65.5	51.4	43.9	39.2	35.2	27.9	23.4	19.1	17.1	16.0	14.6	13.7
150	132.8	81.1	61.5	47.6	40.2	35.7	32.8	26.3	22.0	17.9	16.0	14.9	13.6	12.8
182	129.6	78.0	58.5	44.6	37.4	32.9	30.1	25.6	21.0	17.0	15.2	14.1	12.9	12.1
225	126.7	75.2	55.8	42.0	34.8	30.4	27.7	23.9	19.9	16.1	14.4	13.4	12.2	11.4
273	124.5	73.0	53.7	40.0	32.9	28.6	25.9	22.2	19.1	15.4	13.7	12.8	11.6	10.9
365	121.9	70.5	51.3	37.6	30.6	26.3	23.7	20.0	18.2	14.6	13.0	12.0	10.9	10.3
545	119.3	68.0	48.9	35.2	28.3	24.1	21.5	17.9	16.1	13.6	12.1	11.3	10.2	9.6
730	117.9	66.7	47.6	34.0	27.2	23.0	20.4	16.8	15.0	13.2	11.7	10.8	9.8	9.2

[a] ¢/1000 L = ¢/1000 gal * 0.2642

Table 5. GAC Treatment Cost Estimates by System Size and Bed Life; EBCT = 15 Minutes

GAC Bed Life (days)	Systen Cost (cents/1000 gallons)[a] Design Size (MGD)													
	0.1	0.25	0.5	1	2.5	5	10	15	25	50	75	100	150	200
30	259.0	198.5	174.2	156.2	117.2	91.1	74.7	63.1	55.3	52.7	50.4	48.1	47.3	46.4
45	216.5	157.6	134.4	117.5	102.6	76.6	62.2	51.6	44.7	41.4	38.2	35.7	34.8	34.1
60	195.0	136.8	114.2	97.8	88.9	68.7	55.5	45.4	39.1	32.7	31.7	30.8	28.1	28.0
75	182.0	124.2	101.9	85.9	77.4	63.7	51.2	41.5	35.5	29.6	28.6	27.7	25.2	25.1
90	173.2	115.8	93.7	77.9	69.7	61.4	48.2	38.8	33.1	27.4	24.9	23.3	23.2	21.8
105	166.9	109.7	87.8	72.2	64.1	58.7	46.0	36.8	31.3	25.9	23.4	21.9	21.8	20.4
122	161.6	104.6	82.8	67.3	59.5	54.1	44.1	35.0	29.7	24.5	22.1	20.7	19.9	19.2
150	155.4	98.6	77.0	61.7	54.0	48.8	41.8	32.9	27.9	22.9	20.6	19.2	18.0	17.8
182	150.7	94.0	72.6	57.3	49.8	44.8	40.6	31.2	26.4	21.6	19.4	18.1	16.5	15.6
225	146.4	89.9	68.6	53.4	46.1	41.1	38.2	29.6	25.0	20.4	18.3	17.0	15.5	14.6
273	143.1	86.8	65.5	50.5	43.2	38.3	35.5	28.9	23.9	19.4	17.4	16.2	14.8	13.9
365	139.3	83.1	61.9	47.0	39.8	35.0	32.3	26.5	22.5	18.3	16.3	15.2	13.8	13.0
545	135.5	79.4	58.3	43.5	36.5	31.8	29.1	23.3	21.1	17.0	15.2	14.2	12.9	12.1
730	133.5	77.5	56.5	41.7	34.7	30.1	27.5	21.7	19.5	16.3	14.6	13.6	12.3	11.6

[a] ¢/1000 L = ¢/1000 gal * 0.2642

Table 6. Parameters Used in PTA Analysis

PTA system operated at 70% of design capacity
Maximum tower diameter: 16 feet
Maximum liquid loading: 30 gpm/ft^2
Minimum air gradient: 50 $N/m^2/m$
Influent water temperature: 12°C
Tower packing material: 2 inch plastic saddles
Net efficiency of blower/motor: 35%
Net efficiency of pump/motor: 80%
Tower off-gas velocity: 45 feet/minute
Post-heater gas-phase temperature: 24°C
Gas-phase BPL GAC price: 2.0 $/lb at 20,000 lbs
Gas-phase GAC adsorber bed depth: 3 feet
Off-site GAC regeneration: 75 ¢/lb at 50,000 lbs
GAC regeneration and handling losses: 10%
Capital amortized at 10% over 20 years
Labor and fringe rate: 15 $/hour
Electric rate: 0.08 $/kwh
Contractor profit factor for construction cost: 7%
Special construction sitework factor: 5%
Construction engineering fees factor: 12%
ENR Construction Cost Index (June 1988): 421
Producers Price Index (May 1988): 299

Table 7. PTA Cost Estimates with and without Off-Gas Control for Four System Sizes, Co=100, Ce=5 ug/L

VOC	Henry's Coeff	A:W Ratio	Stripping Factor	Packing Depth (ft)	System Cost (¢/1000 gallons)							
					0.1 mgd		1 mgd		10 mgd		100 mgd	
					PTA	w/Off-Gas	PTA	w/Off-Gas	PTA	w/Off-Gas	PTA	w/Off-Gas
1,2-Dichloroethane	0.023	90	2.07	29	48.9	138.2	21.2	65.5	14.1	48.0	12.2	43.9
p-Dichlorobenzene	0.042	60	2.57	28	45.0	112.0	17.8	42.8	11.6	29.6	9.9	26.6
Benzene	0.06	43	2.58	24.5	41.8	102.7	15.1	36.7	9.5	24.6	7.9	21.7
Trichloroethylene	0.116	30	3.48	22.5	39.1	92.1	13.0	29.8	8.0	18.9	6.6	16.2
1,1,1-Trichloroethane	0.172	25	4.3	21.2	38.3	88.5	12.3	27.7	7.5	17.2	6.1	14.5
Carbon Tetrachloride	0.21	23	4.8	19.2	37.4	85.9	11.7	26.1	7.1	15.8	5.7	13.3
1,1-Dichloroethylene	0.4	19	7.6	15.7	35.6	99.4	10.5	40.0	6.3	28.9	5.1	25.2
Vinyl Chloride	2.4	10	24	14	34.5	1280	9.6	1127	5.7	1021	4.5	931

Note: Water temperature 12°C

Table 8. PTA Cost Estimates with and without Off-Gas Control for Four System Sizes, Co=100, Ce= 1 ug/L

VOC	Henry's Coeff	A:W Ratio	Stripping Factor	Packing Depth (ft)	System Cost (¢/1000 gallons)							
					0.1 mgd		1 mgd		10 mgd		100 mgd	
					PTA	w/Off-Gas	PTA	w/off-Gas	PTA	w/Off-Gas	PTA	w/Off-Gas
1,2-Dichloroethane	.023	100	2.3	43.6	56.6	150.8	27.3	75.7	18.7	56	16.4	51.4
p-Dichlorobenzene	.042	70	3	41.8	51	122.6	23	52.2	15.4	36.2	13.4	32.8
Benzene	.06	50	3	36.7	47.1	111.7	19.4	43.4	12.4	29.7	10.6	26.4
Trichloroethylene	.116	32	3.7	35.1	43.6	97.9	16.5	34	10.3	21.8	8.5	18.8
1,1,1-Trichloroethane	.172	28	4.8	32.4	42.5	94.7	15.5	32	9.6	20.1	7.9	17.2
Carbon Tetrachloride	.21	25	5.3	29.6	41.4	91.3	14.6	29.7	9	18.3	7.3	15.5
1,1-Dichloroethylene	.40	20	8	24.2	39.2	104.7	12.9	43.6	7.8	31.5	6.3	27.4
Vinyl Chloride	2.4	10	24	21.6	37.8	1276	11.8	116	6.9	1011	5.5	922

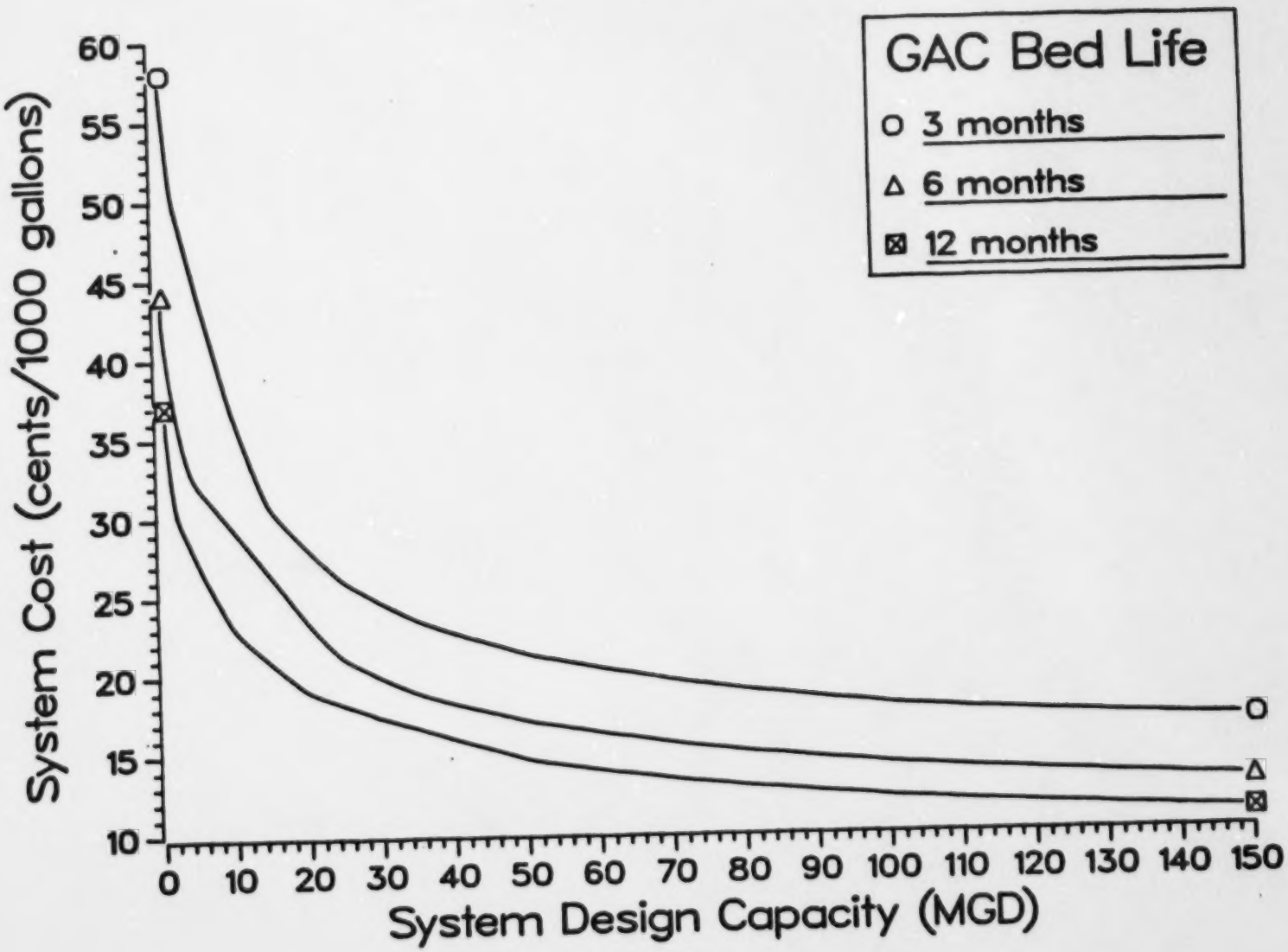

FIGURE 1. GAC TOTAL COST ESTIMATES AS A FUNCTION OF SYSTEM SIZE; EBCT = 10 MINUTES

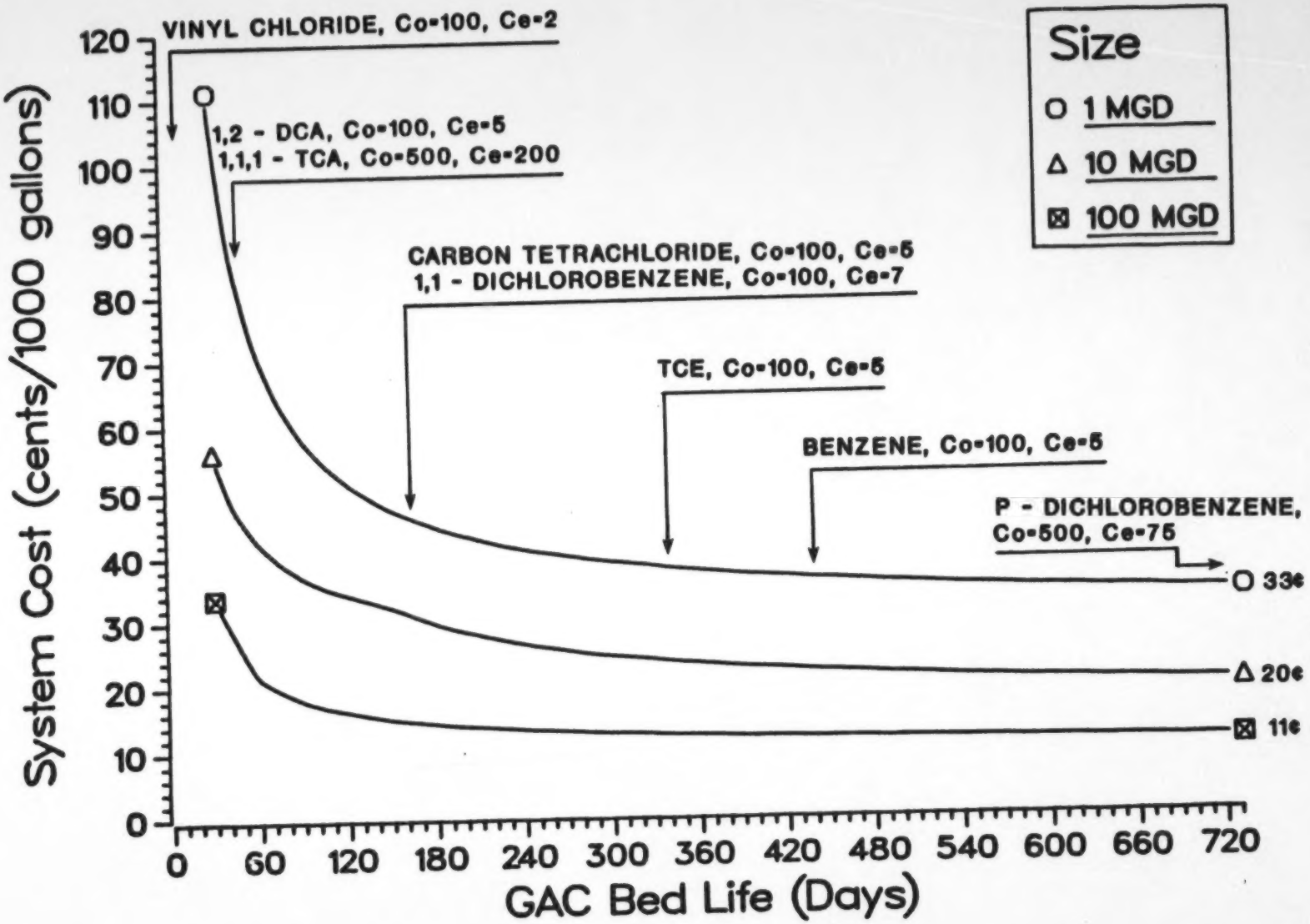

FIGURE 2. TOTAL COST ESTIMATES FOR CONTROLLING SELECTED VOCs BY GAC WITH EBCT = 10 MINUTES

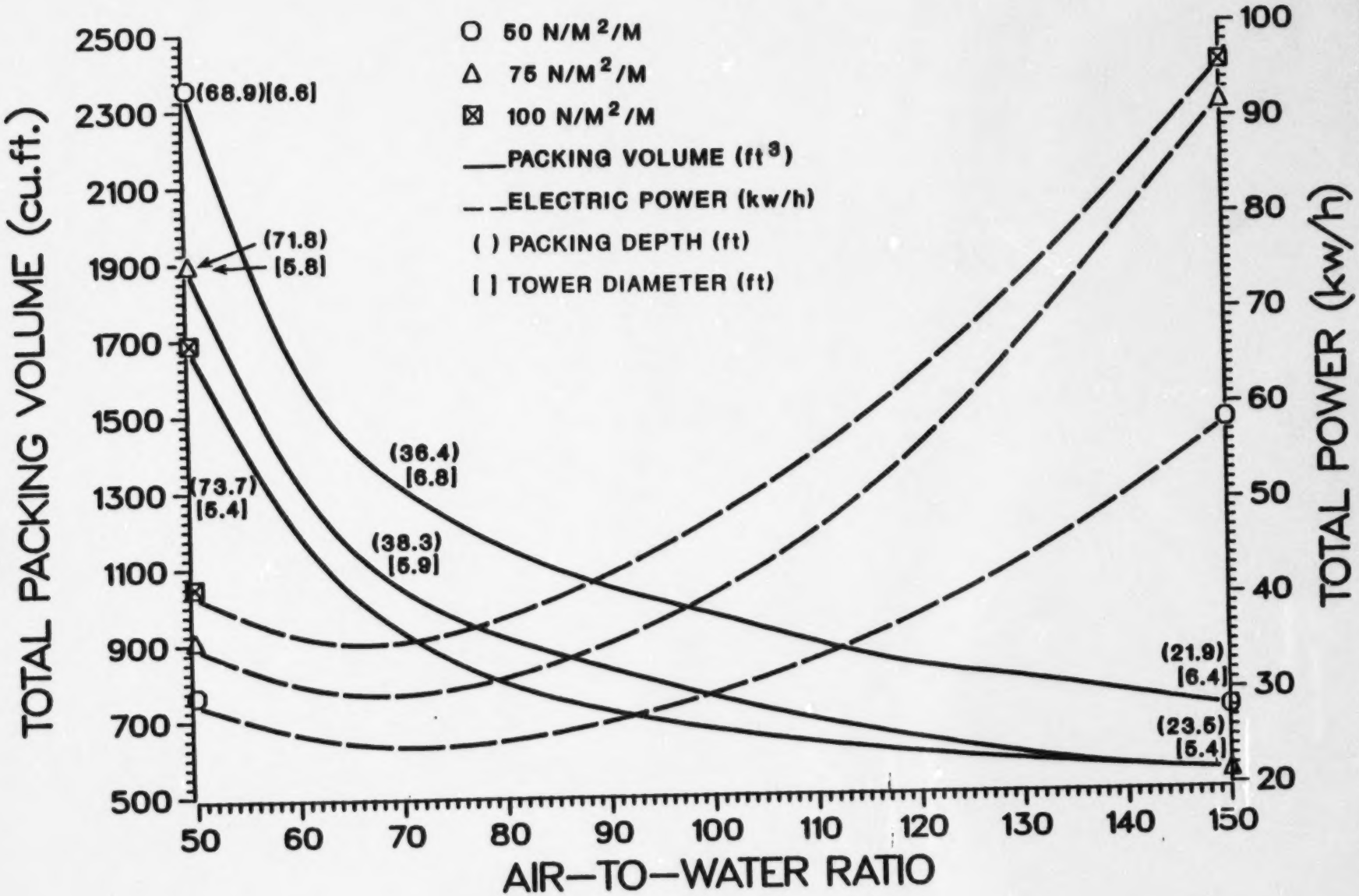

FIGURE 3. DESIGN TRADEOFFS FOR 1 mgd PTA SYSTEM WITH 95% REMOVAL OF 1,2-DCA

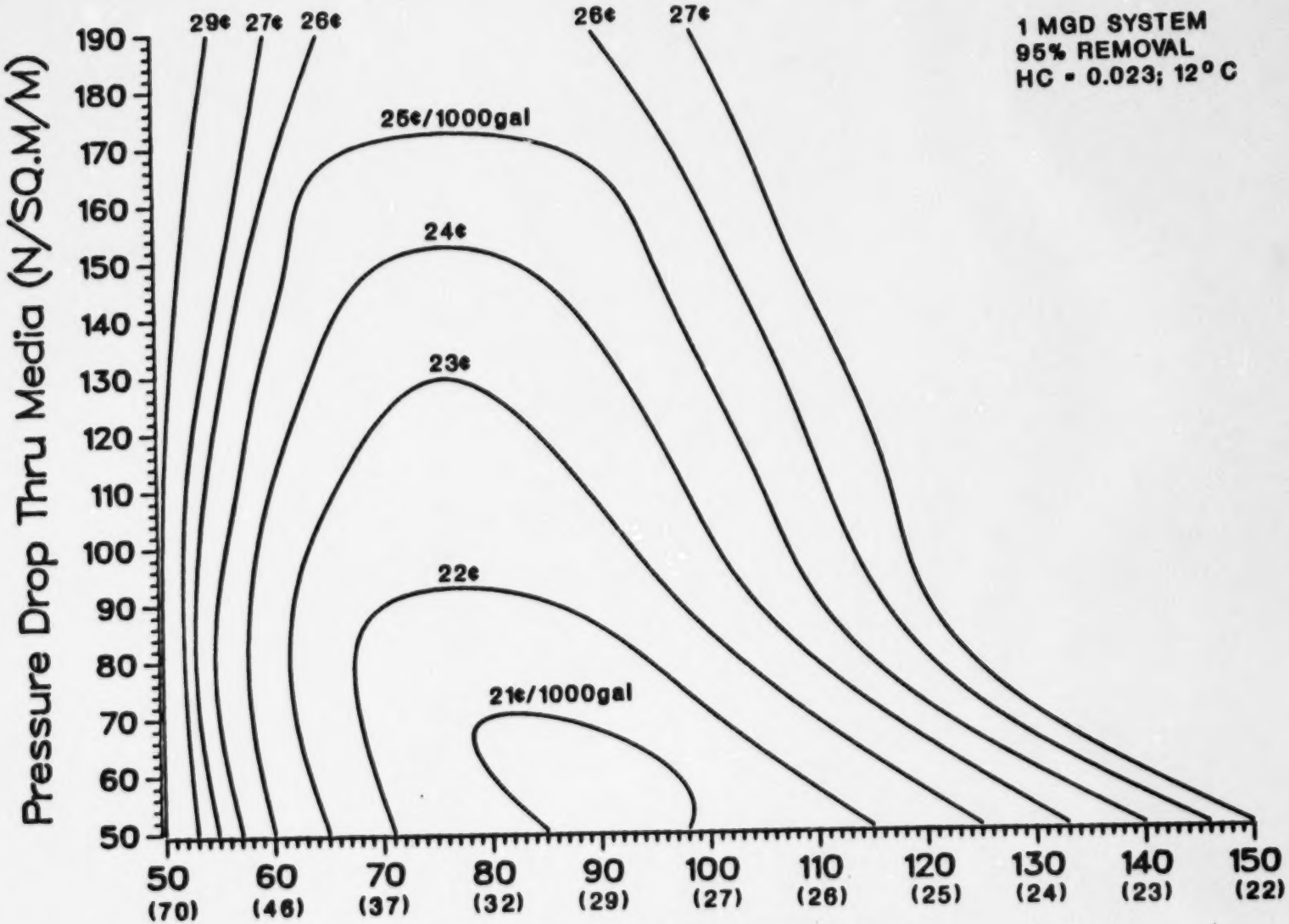

FIGURE 4. COST ESTIMATES FOR CONTROLLING 1,2- DCA BY PACKED TOWER AERATION AS A FUNCTION OF AIR/WATER RATIO AND PRESSURE DROP

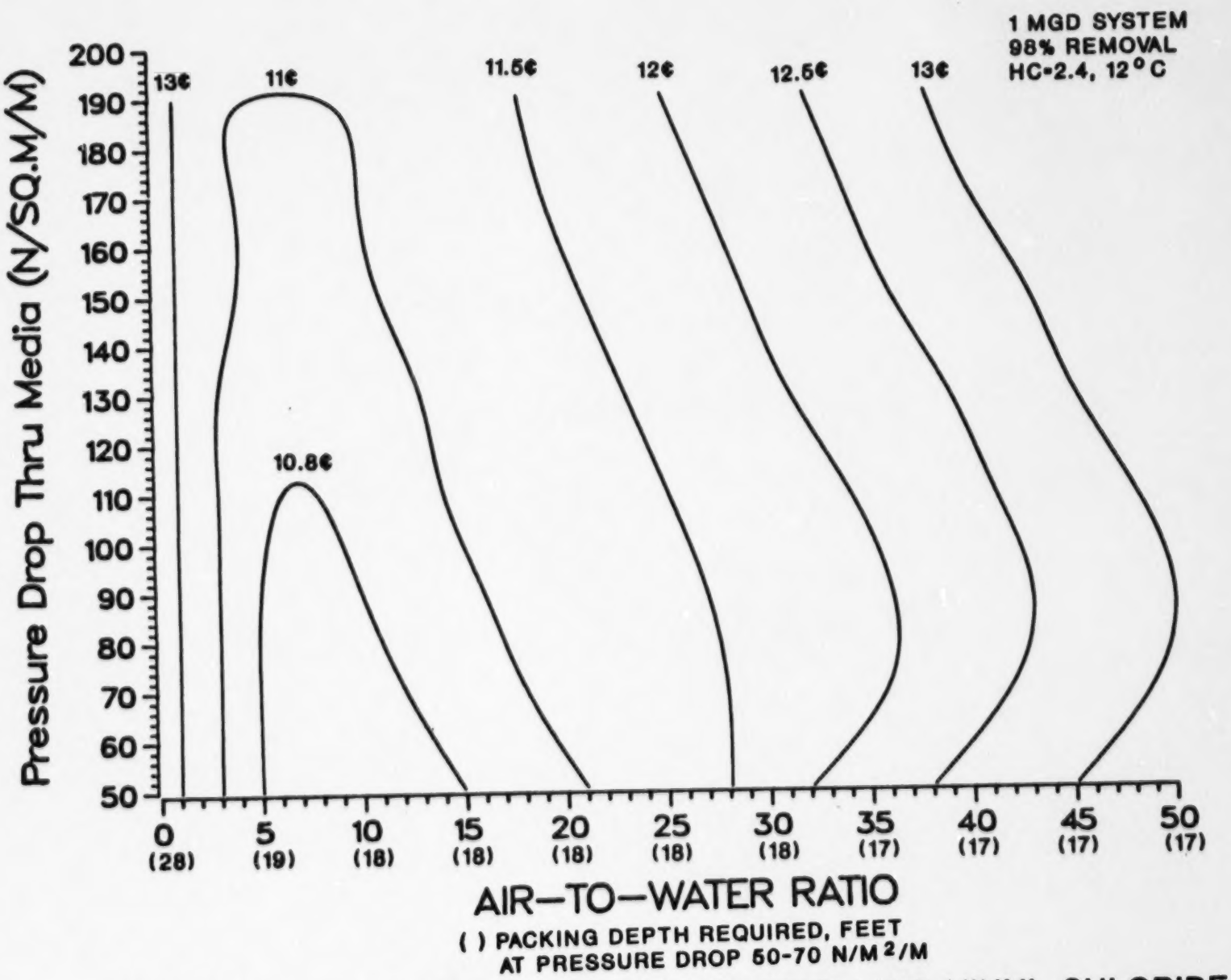

FIGURE 5. COST ESTIMATES FOR CONTROLLING VINYL CHLORIDE BY PTA AS A FUNCTION OF AIR/WATER RATIO AND PRESSURE DROP

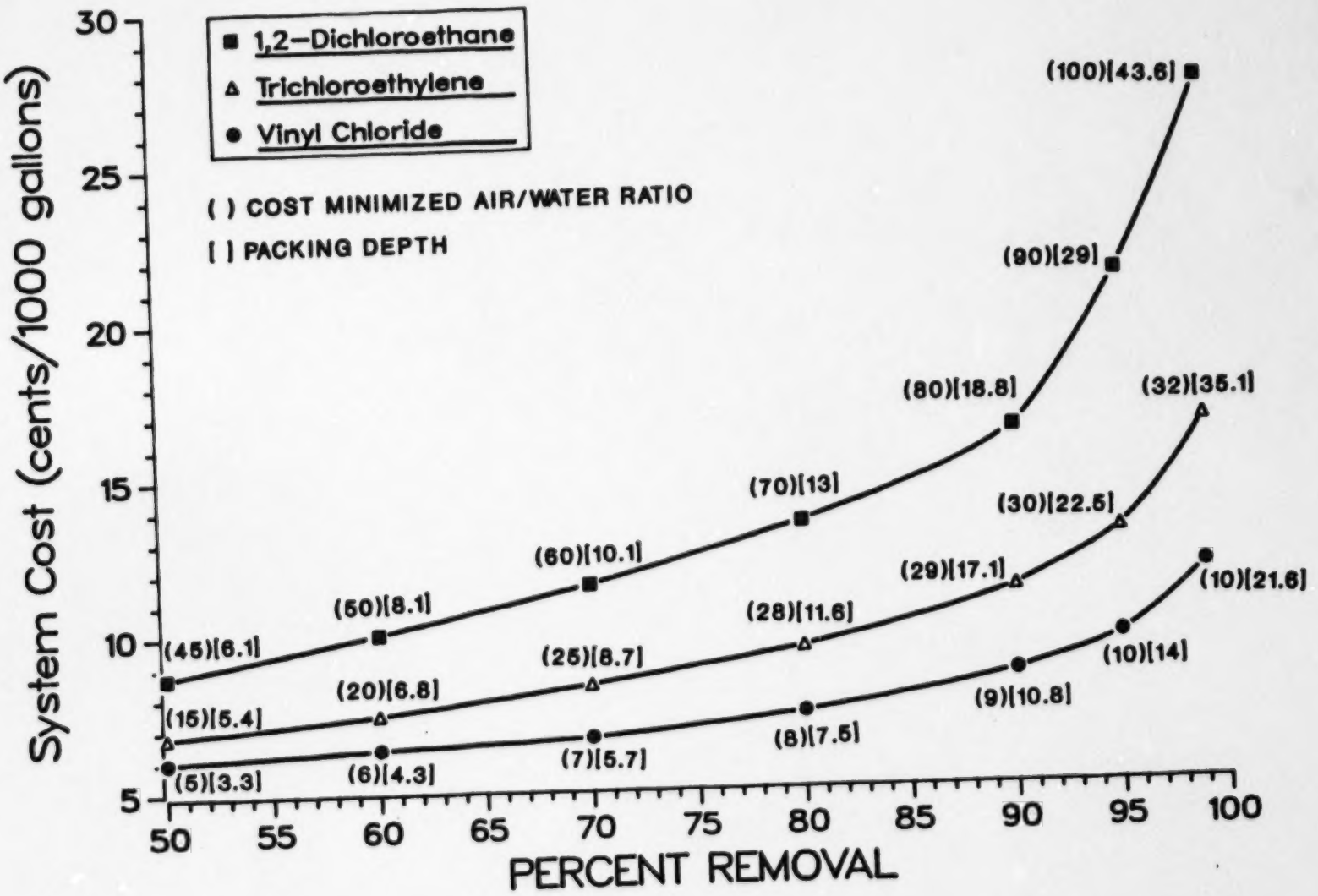

FIGURE 6. COST ESTIMATES FOR CONTROLLING 3 VOCs BY A 1 mgd PTA SYSTEM, AS FUNCTION OF % REMOVAL

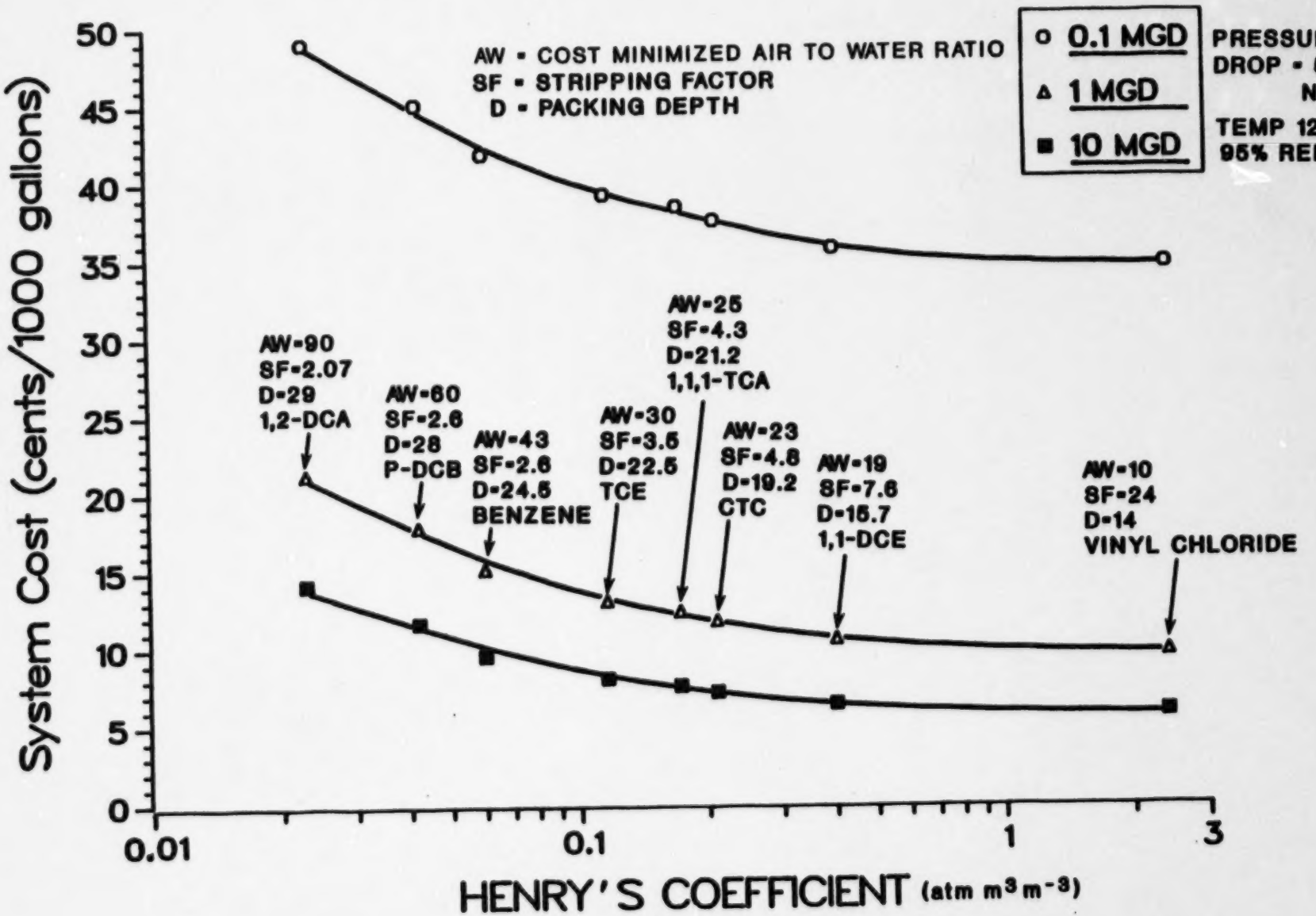

FIGURE 7. COST ESTIMATES FOR CONTROLLING 8 VOCs BY PACKED TOWER AERATION; 95% REMOVAL

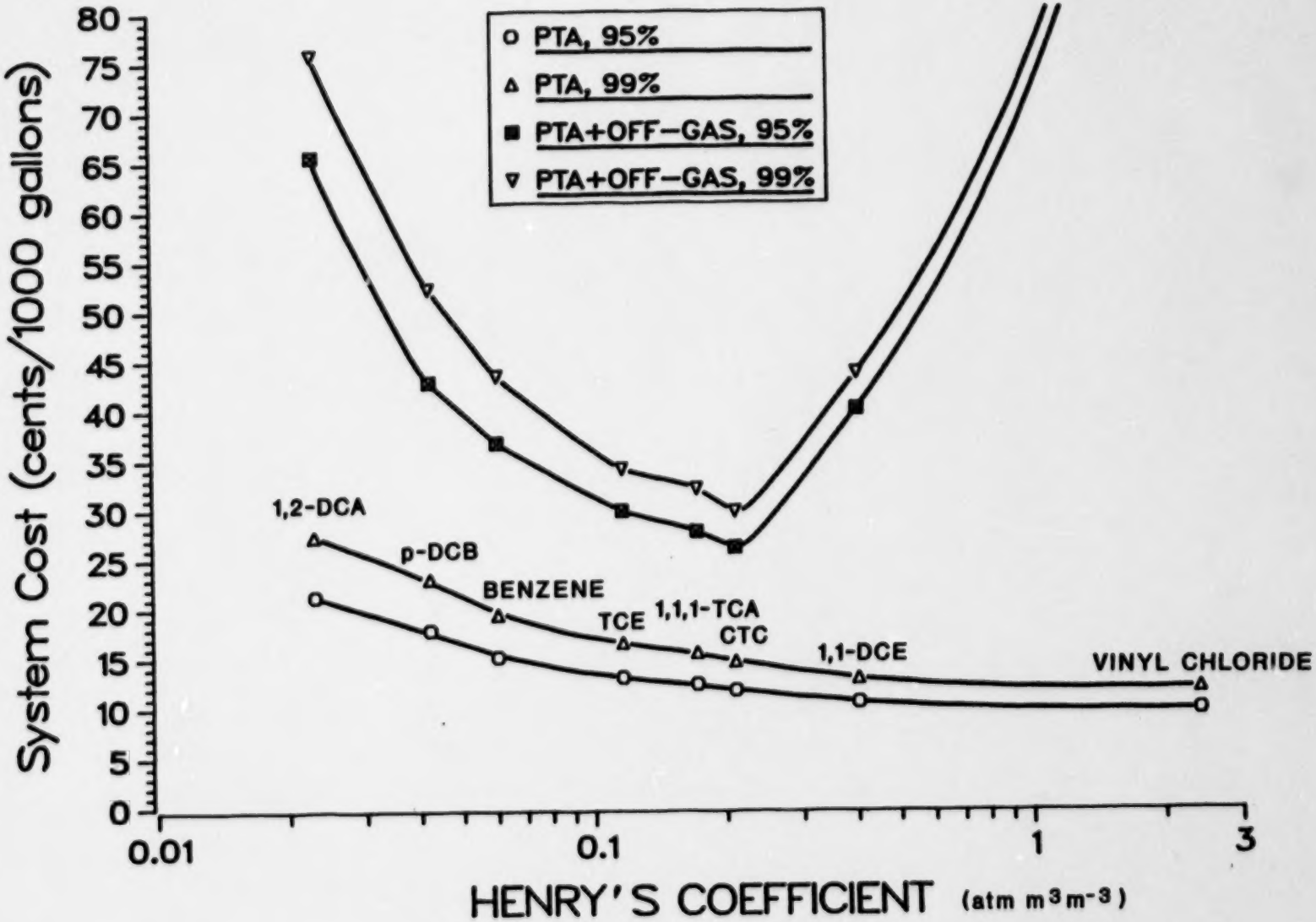

FIGURE 8. COST ESTIMATES FOR CONTROLLING EIGHT VOCs BY PTA FOR A 1 mgd SYSTEM

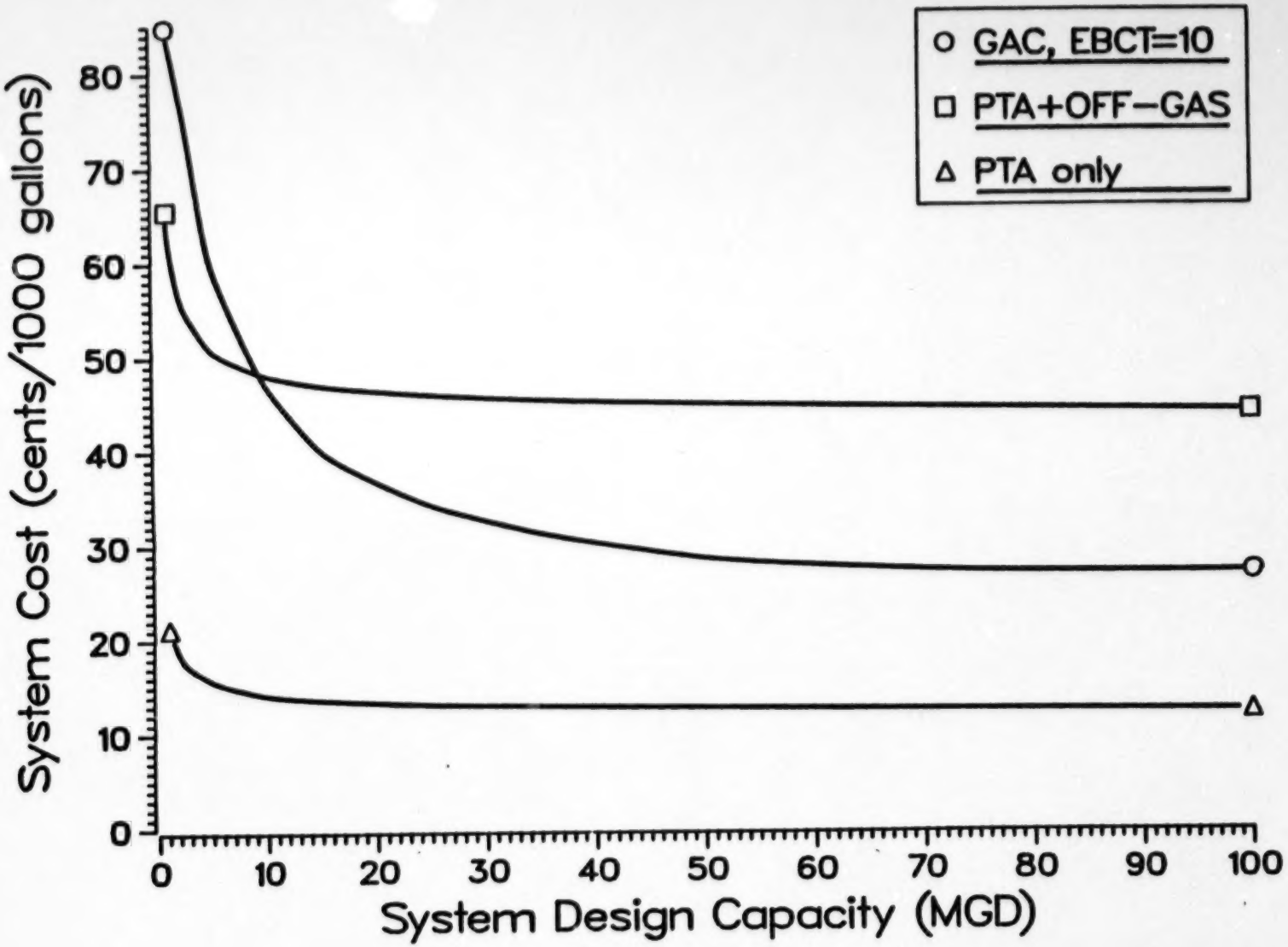

FIGURE 9. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING 1,2 - DICHLOROETHANE Co = 100 ug/L, Ce = 5 ug/L

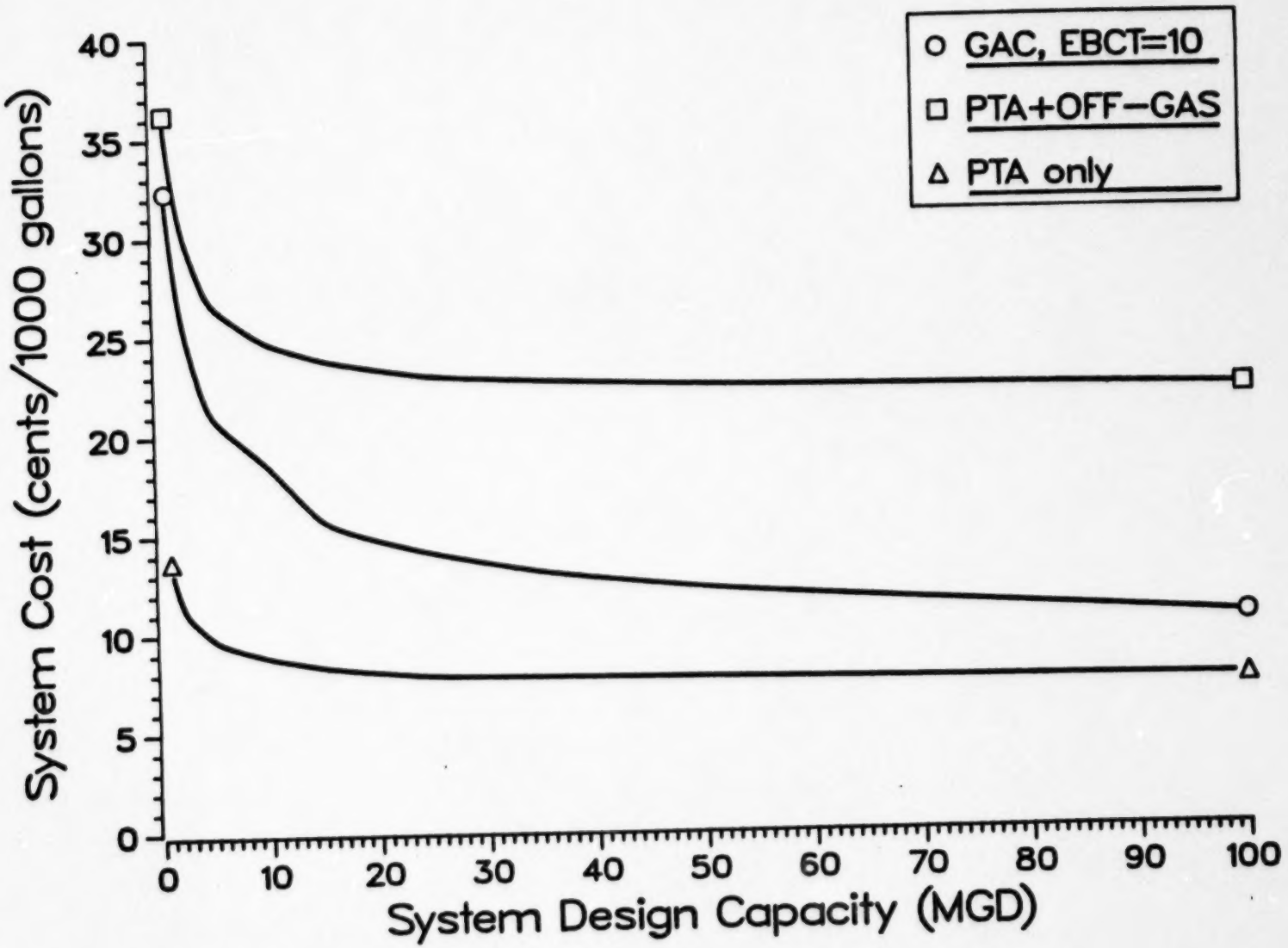

FIGURE 10. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING P - DICHLOROBENZENE Co = 500 ug/L, Ce = 75 ug/L

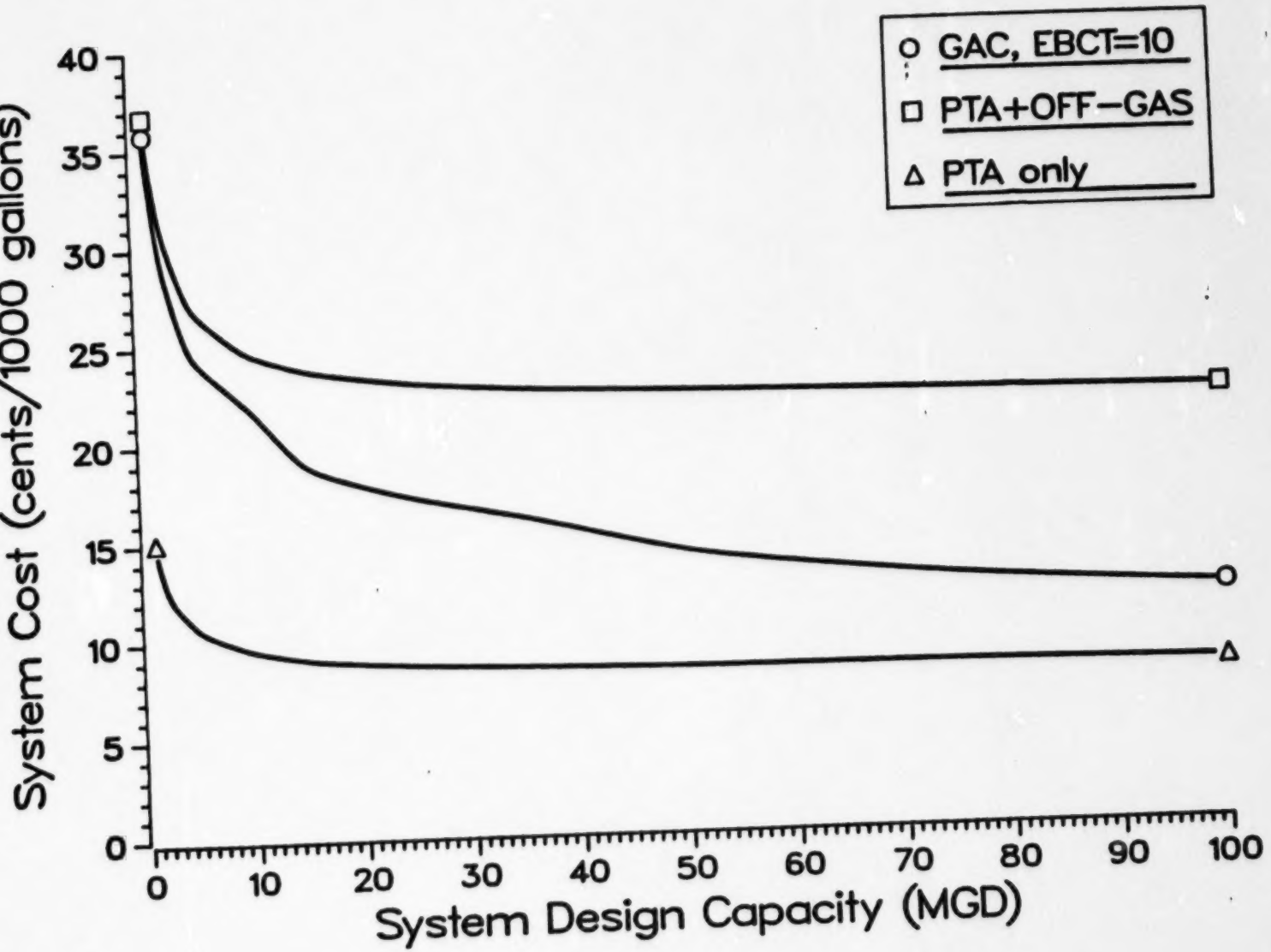

FIGURE 11. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING BENZENE, Co = 100 ug/L, Ce = 5 ug/L

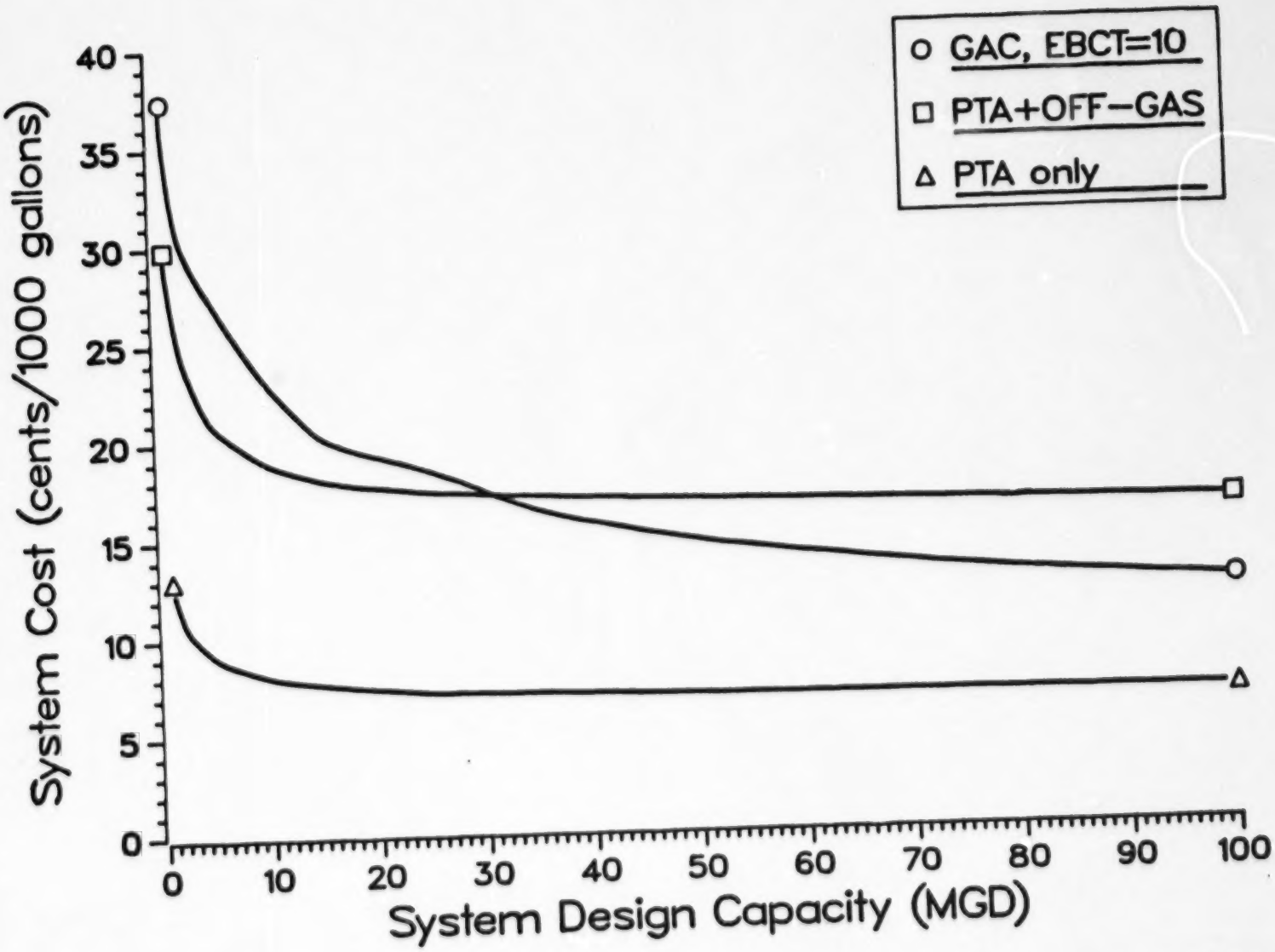

FIGURE 12. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING TRICHLOROETHYLENE, Co = 100 ug/L, Ce = 5 ug/L

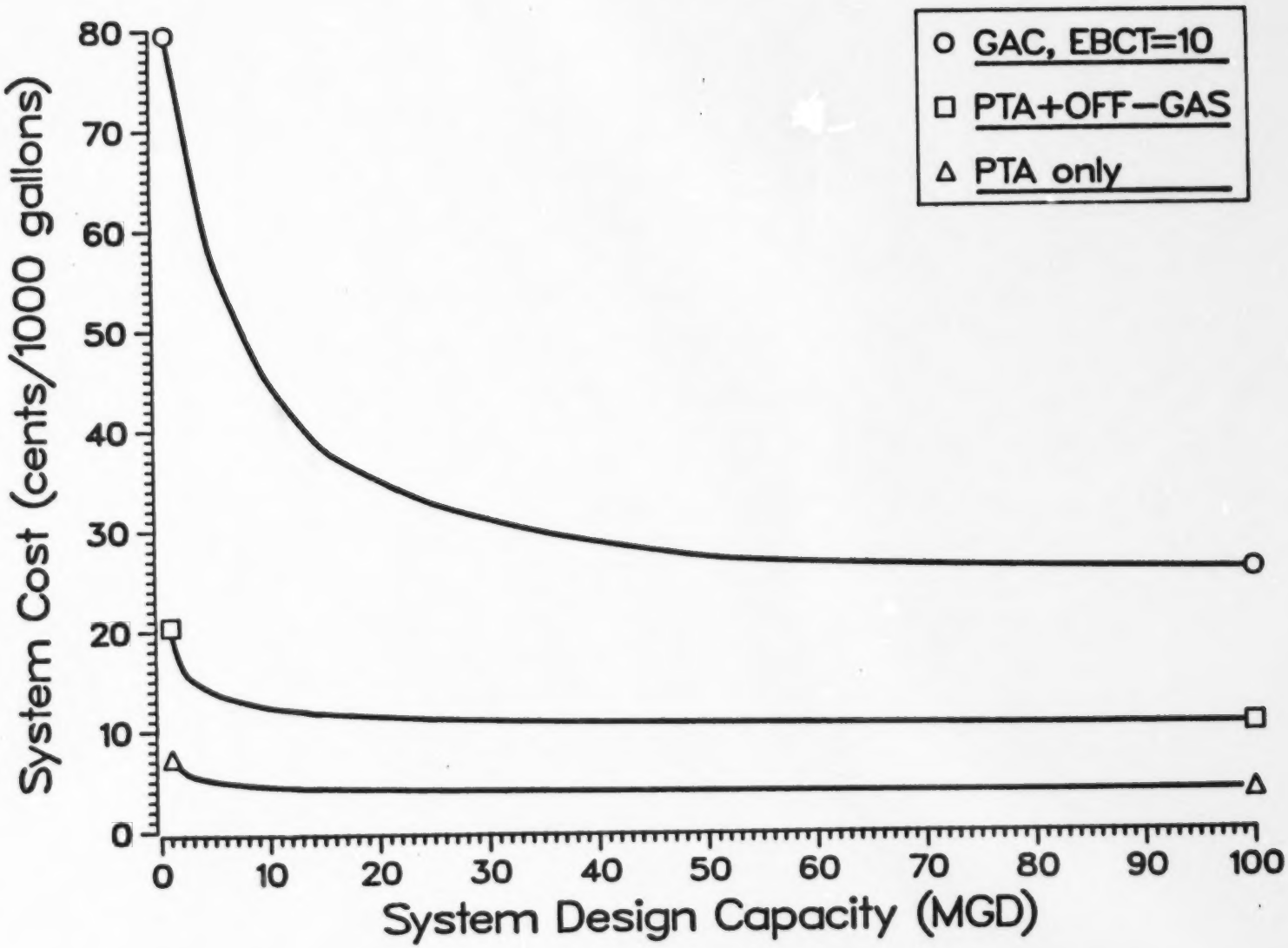

FIGURE 13. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING 1,1,1 - TRICHLOROETHANE, Co = 500 ug/L, Ce = 200 ug/L

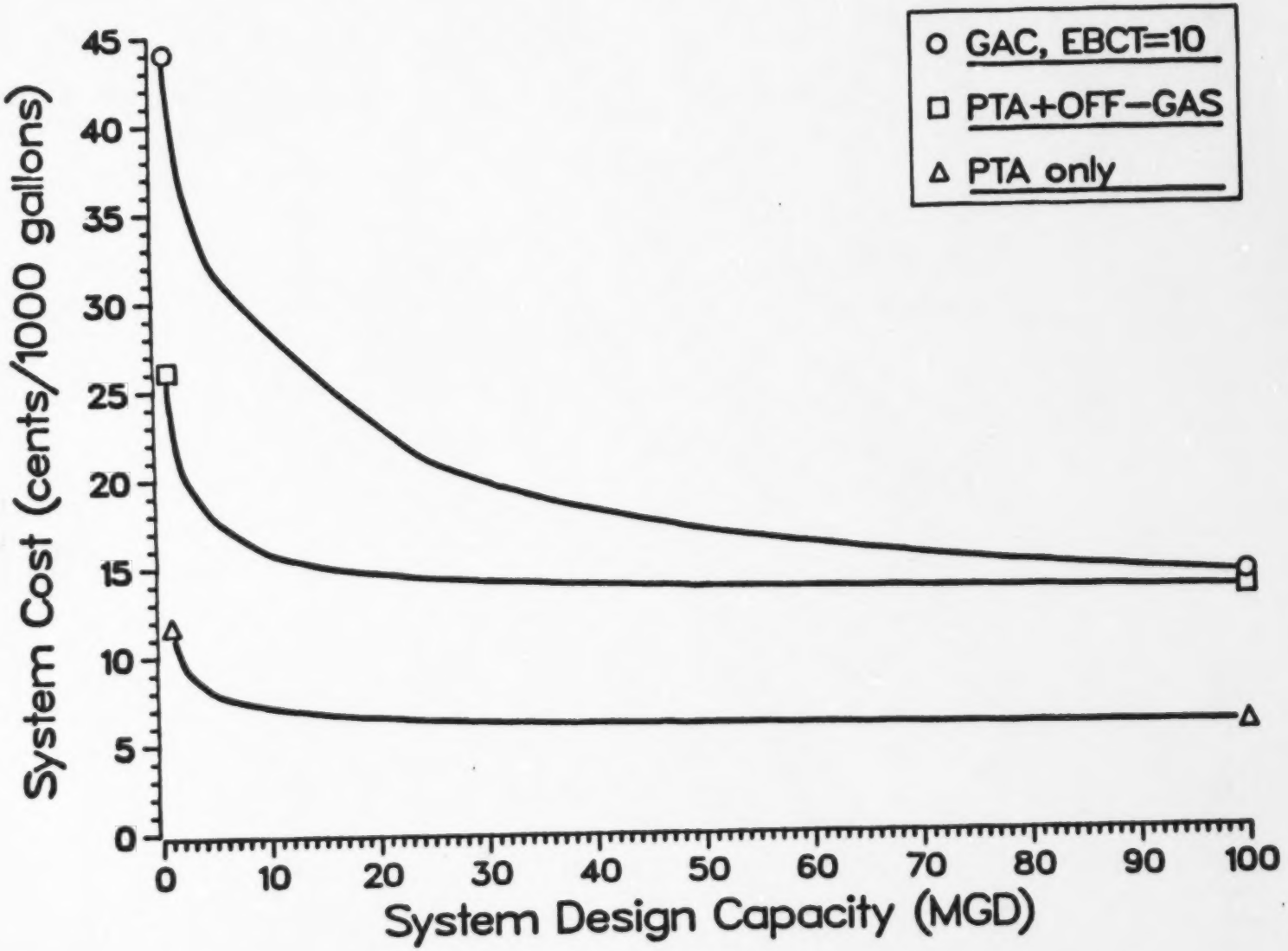

FIGURE 14. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING CARBON TETRACHLORIDE, Co = 100 ug/L, Ce = 5 ug/L

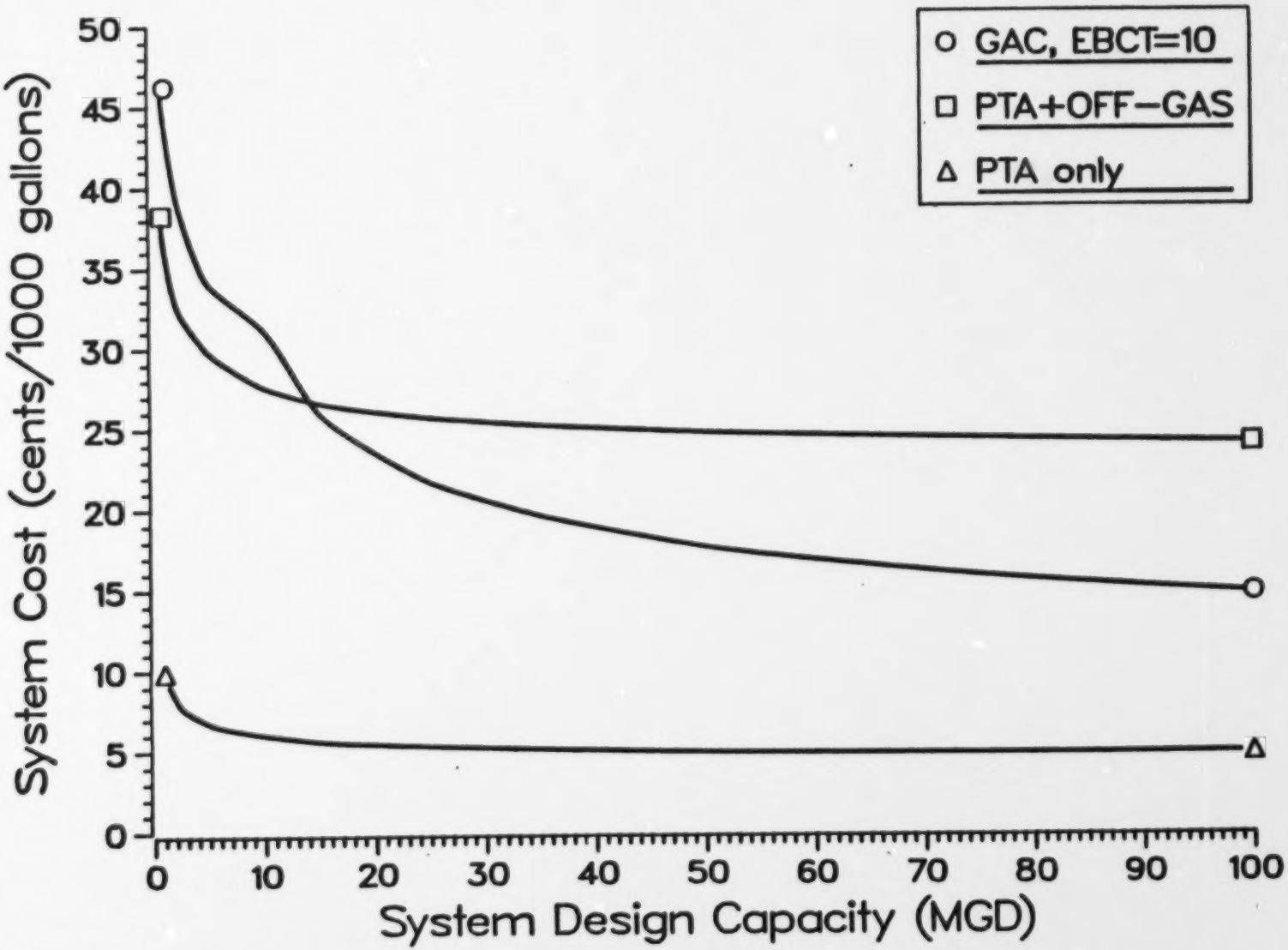

FIGURE 15. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING 1,1 - DICHLOROETHYLENE, Co = 100 ug/L, Ce = 7 ug/L

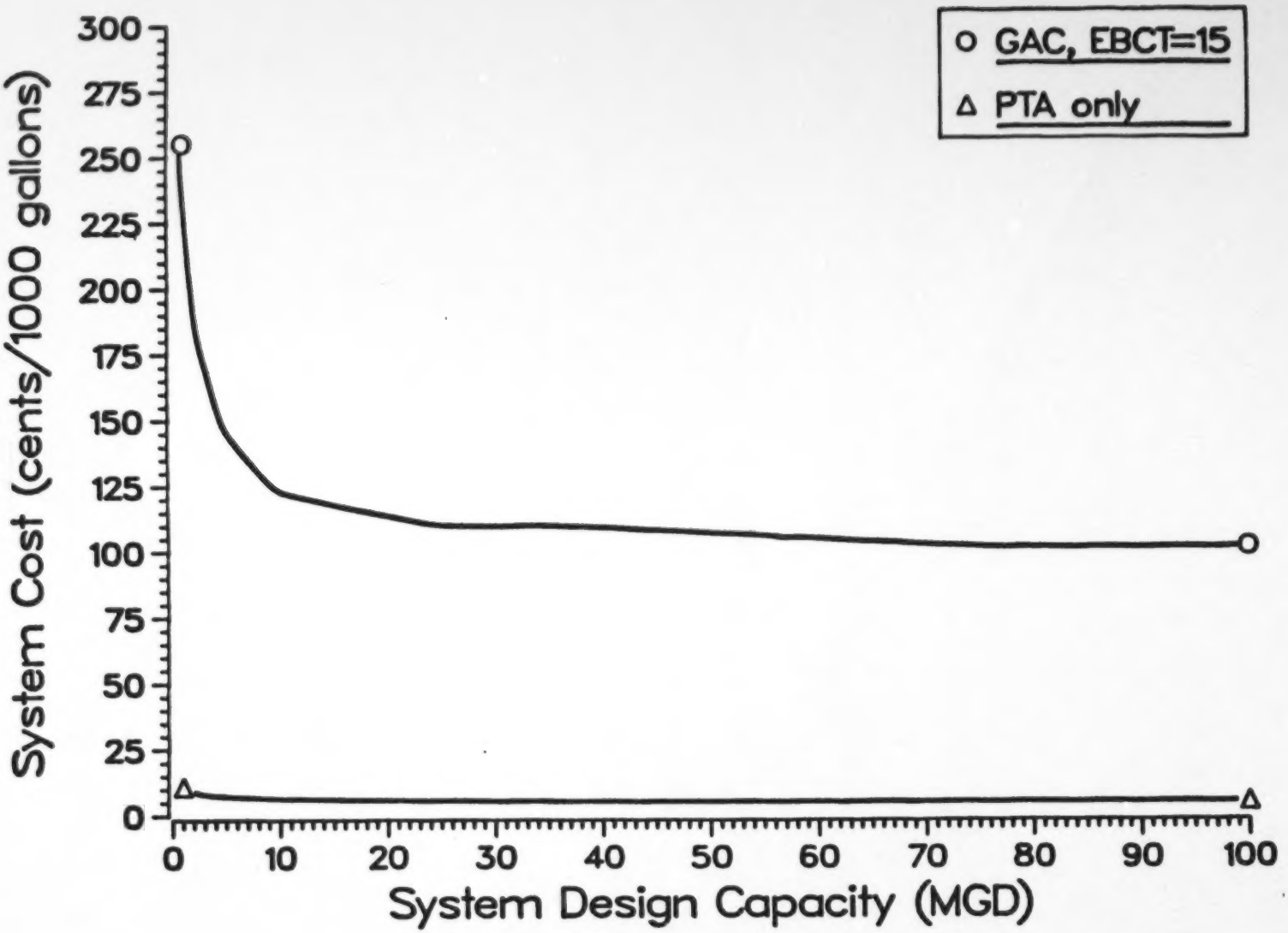

FIGURE 16. LIQUID - PHASE GAC VS. PTA FOR CONTROLLING VINYL CHLORIDE, Co = 100 ug/L, Ce = 2 ug/L